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Volume VI

PRODUCTION OF JET FUELS FROM COAL DERIVED LIQUIDS

VOL VI - Preliminary Analysis of Upgrading Alternatives For The Great Plains Liquid By-Product Streams

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This technical report has been reviewed and is approved for publication.

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Amoco and Lummus Crest have developed seven cases for upgrading by-product liquids from the Great Plains Coal Gasification plant to jet fuels, and in several of the cases, saleable chemicals in addition to jet fuels.

The analysis shows that the various grades of jet fuel can be produced from the Great Plains tar oil, but not economically. However, the phenolic and naphtha streams do have the potential to significantly increase (on the order of $10-15 million/year) the net revenues at Great Plains by producing chemicals, especially cresylic acid, cresol, and xylenol. The amount of these chemicals, which can be marketed, is a concern, but profits can be generated even when oxygenated chemical sales are limited to 10 percent of the U.S. market. Another concern is that while commercial processes exist to extract phenolic mixtures, these processes have not been demonstrated with the Great Plains phenolic stream.

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SUMMARY

Amoco and Lummus Crest are contracted with DOE to develop an upgrading scheme for the liquid by-products (tar oil, phenols, and naphtha) produced by the Great Plains gasifiers. These streams are currently burned in the utility boilers and steam superheaters in the Great Plains plant. Task 1 of the contract is complete and the results are reported here. The objective of this task is to develop preliminary economics for seven different upgrading schemes with product slates ranging from maximum jet fuel (JP-4, JP-8, and JP-8X) production to chemical production (cresylic acid, phenol, cresol, xylenol, benzene, toluene, and xylene). A linear program has been developed to evaluate the various upgrading alternatives.

The analysis shows that the various grades of jet fuel can be produced from the Great Plains tar oil, but not economically. However, the phenolic and naphtha streams do have the potential to significantly increase (on the order of $10-15 million/year) the net revenues at Great Plains by producing chemicals, especially cresylic acid, cresol, and xylenol. The amount of these chemicals, which can be marketed, is a concern, but profits can be generated even when oxygenated chemical sales are limited to ten percent of the U. S. market. Another concern is that while commercial processes exist to extract phenolic mixtures, these processes have not been demonstrated with the Great Plains phenolic stream.

The revenues from chemical sales are sufficient to subsidize the production of jet fuel from the tar oil stream. The economics for these cases are very sensitive to the cost of the replacement fuel for the utility boilers. Replacement fuel costs cannot exceed approximately $3.00/million Btu if a ten percent real rate of return is to be realized on the capital invested in the upgrading plant.

It should be stressed that these results are preliminary. Numerous assumptions, which need to be verified in subsequent tasks or additional projects, have been made in arriving at the above conclusions.

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FOREWORD

In September 1986, the Fuels Branch of the Aero Propulsion Laboratory at Wright-Patterson Air Force Base, Ohio, commenced an investigation of the potential for production of jet fuel from the liquid by-product streams produced by the gasification of lignite at the Great Plains Gasification Plant located in Buelah, North Dakota. Funding was provided to the Department of Energy (DOE) Pittsburgh Energy Technology Center (PETC) to administer the experimental portion of this effort. This report details the effort of AMOCO Oil Company, who as a contractor to DOE (DOE Contract Number DE-AC22-87PC90015), conducted a preliminary analysis of upgrading alternatives for the production of turbine fuels from the Great Plains liquid by-product streams. DOE/PETC was funded through Military Interdepartmental Purchase Request (MIPR) FY1455-86-N0657. Mr. William E. Harrison III was the Air Force Program Manager, Mr. Gary Stiegel was the DOE/PETC Program Manager, and Drs. Bruce Fleming and Mark Furlong were the AMOCO Program Managers.

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SECTION I

INTRODUCTION

The Great Plains Coal Gasification Plant in Beulah, North Dakota, produces about 145 MM SCF/D of substitute natural gas (SNG) from lignite. The plant also produces three liquid by-products: about 2,900 B/D of tar oil, 830 B/D of crude phenols, and 650 B/D of naphtha. These liquids are all products from the devolatilization of lignite in the Lurgi gasifiers. Currently, the by-products are burned in the plant's boilers and superheaters to produce steam. The economic viability of the plant might be improved by producing marketable products, rather than steam, from these by-product liquids. To this end, Amoco and Lummus Crest, under a contract with the United States Department of Energy, are investigating the technical and economic feasibility of converting the by-product liquids to jet fuels and other saleable products. Jet fuels are of particular interest because of the close proximity of Great Plains to several U. S. Air Force bases.

SECTION II

PROJECT OVERVIEW

As shown in Figure 1, this project is divided into five major tasks: Process Concept Definition, Bench Scale Testing, Pilot Plant Testing, Preliminary Process Design and Economics, and Production Run Recommendation. The results of the first task are reported here.

The first task, Process Concept Definition, includes three subtasks: Liquid By-product Analysis, Process Modelling and Design, and Economic Modelling. The first of these subtasks (1.1), By-Product Analysis, involves analytical characterizations of samples of each by-product taken at six-week intervals. The results from this program, which provide an indication of the average quality of each stream and the variability of that quality over time, are an important input to the second subtask (1.2), Process Modelling and Design. Other inputs to the second subtask include limited experimental processing data on the Great Plains by-products by the Western Research Institute (WRI),[(1)] Amoco's petroleum refining process models and linear programming technology, Lummus' process simulation and design programs and a market analysis of by-products from Great Plains developed by Sinor Consultants.[(2)] In addition, throughout Task 1, ANG Coal Gasification Company provided valuable input and advice on all fronts. The major objective of subtask 1.2 is to produce seven conceptual designs and associated capital and operating costs for facilities to refine the Great Plains by-products. These seven designs are listed in Figure 2. They include designs for maximizing production of each grade of jet fuel (JP-4, JP-8, JP-8X), designs for profitable schemes which produce the various jet fuel grades, and a scheme for maximizing profits. In subtask 1.3 the results generated by Amoco and Lummus are subjected to economic analysis.

The two products from Task 1 are the design and economic results for each of the seven designs and a plan for bench scale testing (Task 2) to confirm any assumptions made in Task 1. Based on the design and economic results from Task 1 and the experimental results from Task 2, the Department of Energy and the Department of Defense will decide on a preferred processing scheme for the

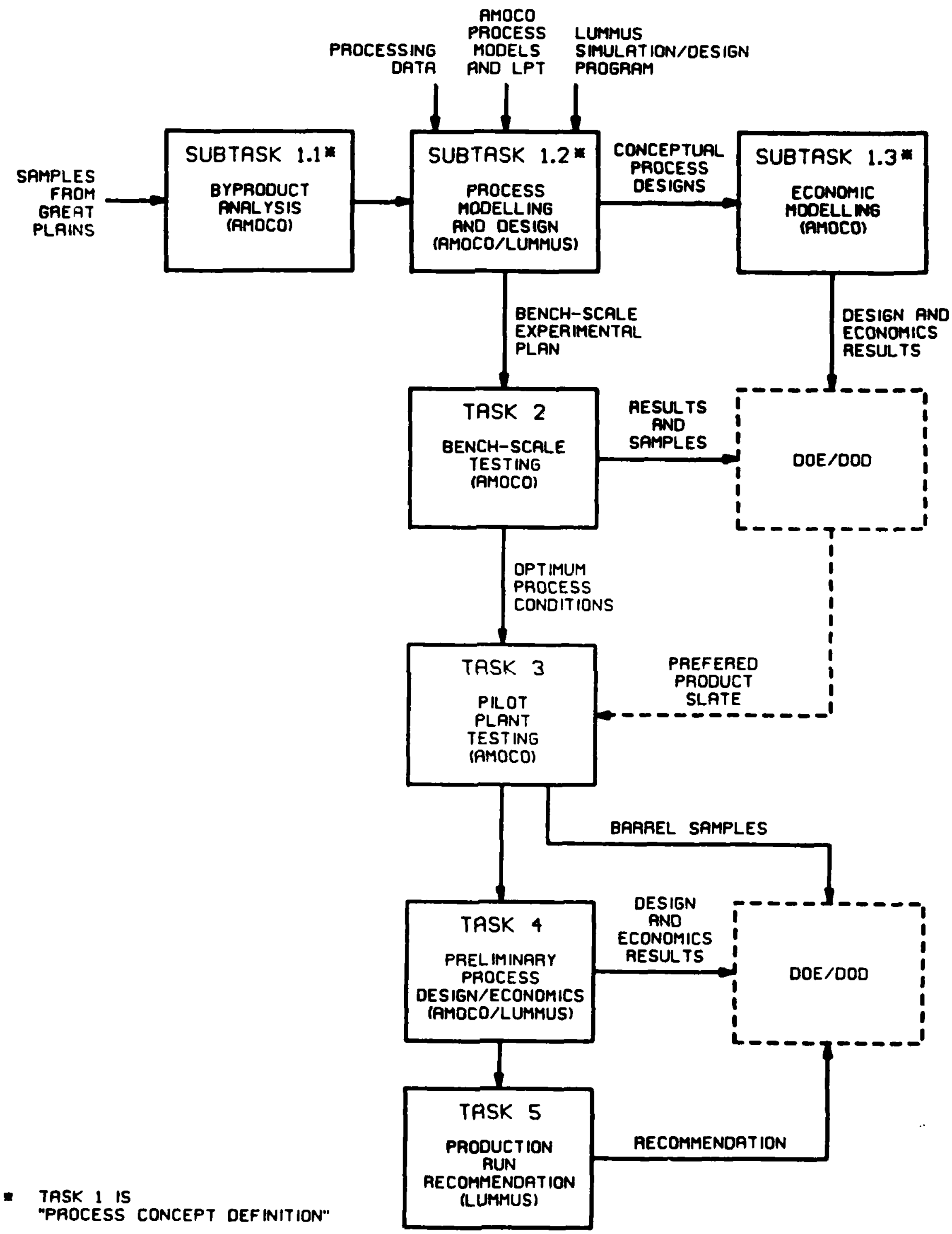

FIGURE 1
PRODUCTION OF JET FUEL FROM COAL DERIVED LIQUIDS: AMOCO/LUMMUS ACTIVITIES

THE FOLLOWING DESIGN CASES WILL RESULT FROM ACTIVITIES IN THIS SUBTASK:

CASE	DESCRIPTION
1	MAXIMUM JP-4 PRODUCTION.
2	PROFITABLE PRODUCT SLATE INCLUDING JP-4.
3	MAXIMUM JP-8 PRODUCTION.
4	PROFITABLE PRODUCT SLATE INCLUDING JP-8.
5	MAXIMUM JP-8X PRODUCTION.
6	PROFITABLE PRODUCT SLATE INCLUDING JP-8X.
7	MAXIMUM PROFIT.

NOTE: CASES 2, 4, 6, AND 7 REQUIRE LINEAR PROGRAMMING TECHNOLOGY.

FIGURE 2

SUBTASK 1.2: PROCESS MODELING AND DESIGN CASE SUMMARY

Great Plains liquids. Amoco will then carry out pilot plant testing (Task 3) to confirm the process design and to provide barrel quantities of product for testing by the government. The pilot plant results will be used by Amoco and Lummus to develop a preliminary process design (Task 4) for a plant to upgrade the liquid by-products at Great Plains. Finally, in Task 5, Lummus will locate existing facilities where the processing scheme can be carried out on a scale sufficient to provide jet fuel for aircraft testing.

SECTION III

SOURCE OF BY-PRODUCT LIQUIDS

Tar oil, crude phenols, and naphtha are produced at the Great Plains Gasification Plant; a schematic of the plant is shown in Figure 3. The plant currently produces about 145 MMSCFD of synthetic natural gas (SNG) from North Dakota lignite. The SNG is composed almost entirely of methane, which is derived mostly from synthesis gas (H_2 + CO) produced in the Lurgi Mark IV gasifiers and methanated in downstream reactors. Some methane is produced by devolatilization of the coal in the gasifiers. The liquid by-products (tar oil, phenolics, and naphtha) are produced during lignite devolatilization in the gasifiers.

The tar oil and phenolics are condensed from the product gas along with water vapor to form a gas liquor. This condensation takes place in heat exchangers located in the gasifier quench, shift converter, gas cooling, and Rectisol units. The liquor is routed to the gas liquor separation unit, where the tar oil is recovered by gravity separation. The heaviest portion of the tar oil, which contains about 20 percent coal dust, is recycled to the gasifiers. The recycle rate of this "dusty tar" is about 1800 B/D. The remaining tar oil, which contains 2-6 percent dust, is produced at a rate of 2900 B/D. The phenolics are recovered from the gas liquor by extraction with isopropyl ether in the Phenolsolvan unit. The resulting crude phenol stream, which is produced at a rate of about 830 B/D is composed mostly of phenol, cresol, and xylenol, with the remainder being water and neutral oils. The naphtha is condensed from the gasifier raw gas by contacting the stream with cold methanol in the Rectisol unit. The naphtha is produced at a rate of 650 B/D.

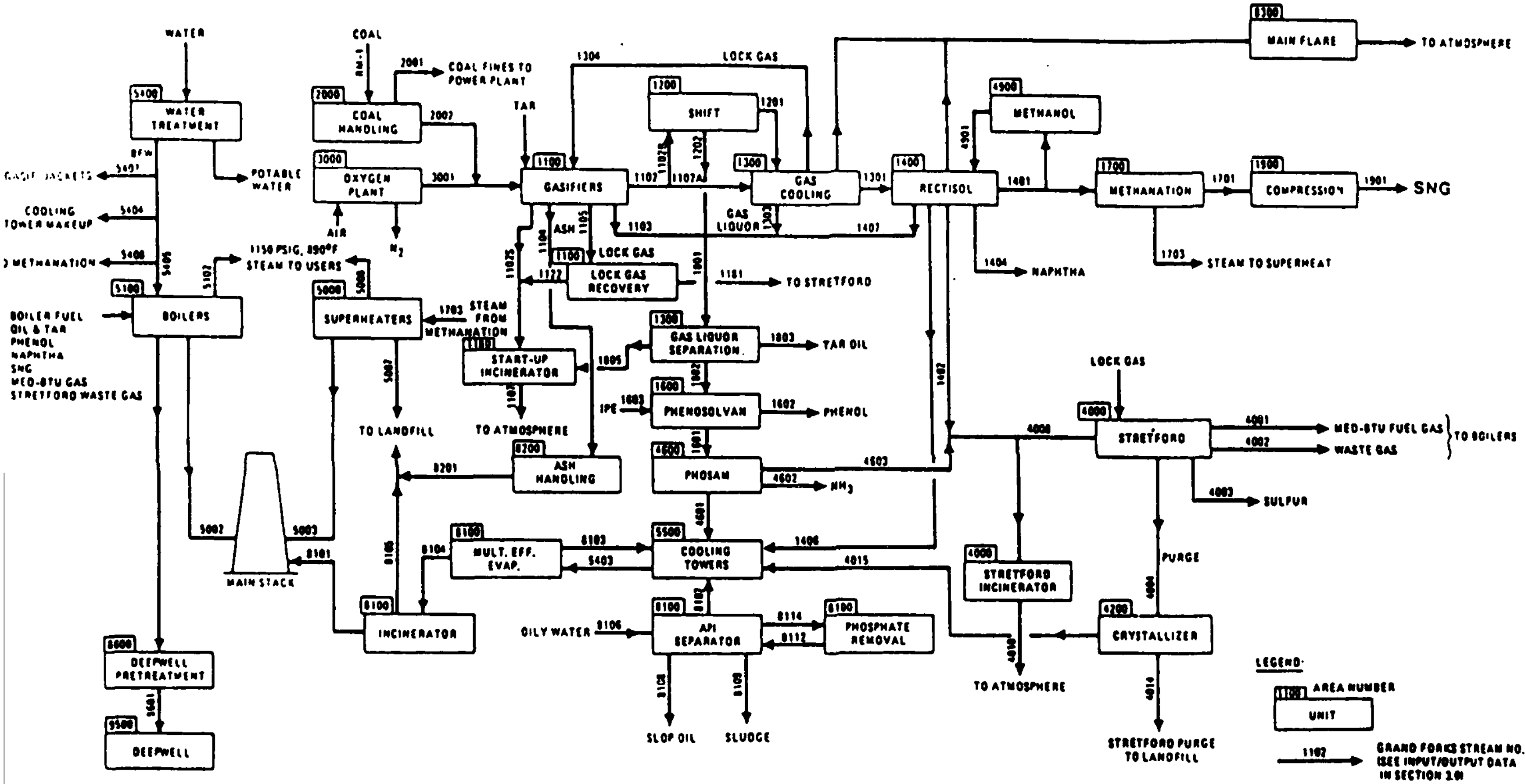

FIGURE 3. GREAT PLAINS GASIFICATION PROCESS BLOCK FLOW DIAGRAM

SECTION IV

TASK 1 RESULTS

1. Subtask 1.1 By-Product Analysis

Each of the three by-product streams from the Great Plains Coal Gasification Plant have been sampled at six to eight week intervals since May of 1987 to monitor seasonal and operational variations in quality and to characterize them as feedstocks for the processes under consideration. This program will continue until mid-1988. These data, with hydrotreating kinetics results from Task 2, will be utilized in Task 4 to prepare a preliminary design of the proposed commercial upgrading/refining facility, integrated into the existing Great Plains plant.

a. Tar Oil

The tar oil stream is the most viable of the by-product streams as a feedstock for the production of jet fuel. ASTM distillations and analyses characterizing this stream are shown in Figure 4 and Table 1. All distillation results have been obtained according to ASTM Method D-2887 with an aromatic standard, except for one D-86, and one D-1160 corrected to D-86. D-86 and D-1160 results are reported on a volume basis and D-2887 on a weight basis. While volume results cannot be rigorously converted to weight without a density profile of the material, the distillations show fairly constant composition among the samples tested. About 2-6 percent of the material boils at less than 300°F and 8-20 percent boils above 800°F on a weight basis.

Previous analyses at the University of North Dakota Energy Research Center (UNDERC)(3) by Wilson and co-workers showed the tar oil contained 90 to 95 percent aromatics, the remainder being paraffins. WRI's(1) hydrogenation data show that significant amounts of jet fuel can be made from this stream, but only at severe processing conditions (2,000 psig) and high hydrogen

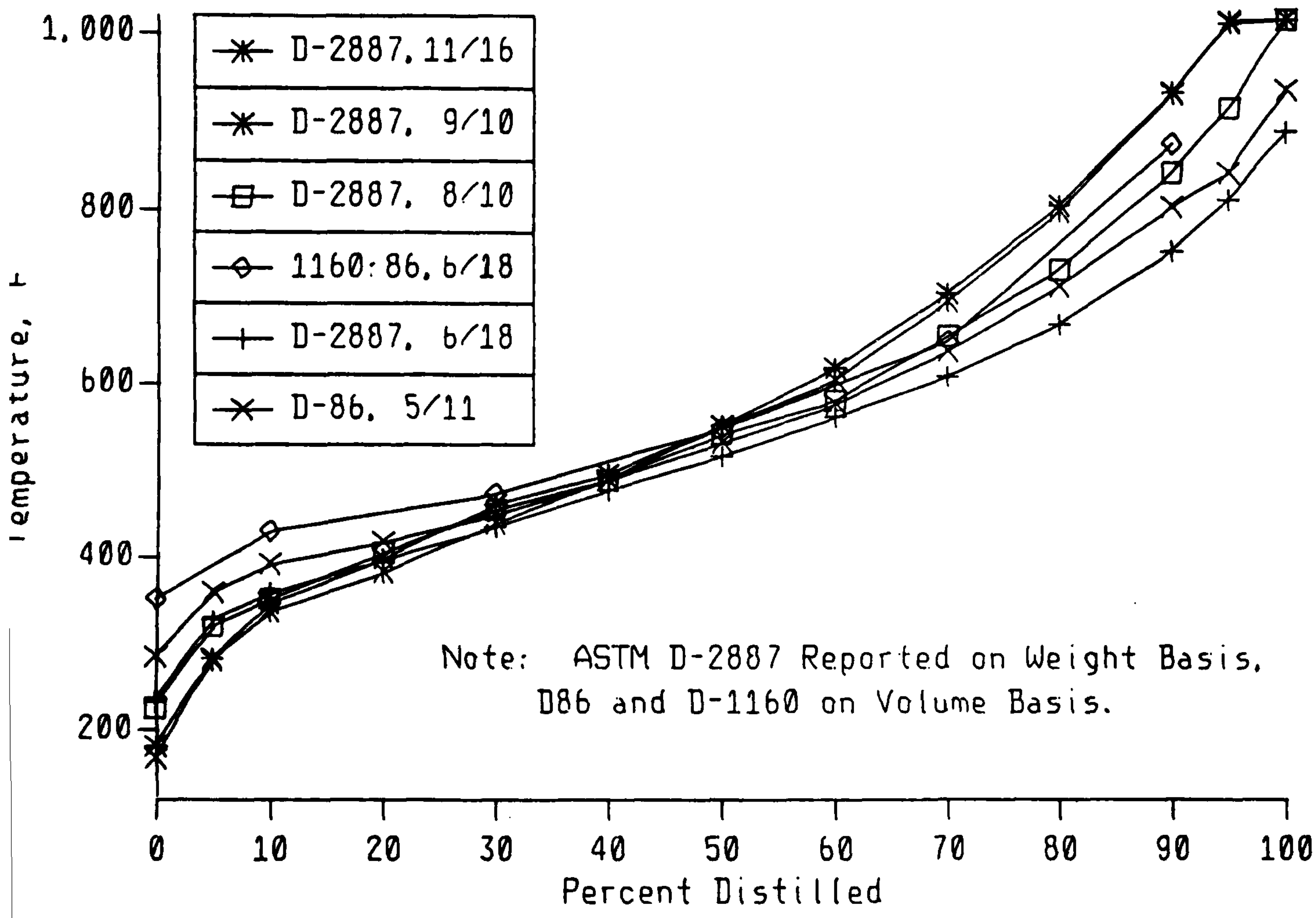

FIGURE 4
GREAT PLAINS COAL LIQUIDS UPGRADING
TAR OIL DISTILLATION RESULTS

TABLE 1

GREAT PLAINS COAL LIQUIDS UPGRADING

FEEDSTOCK CHARACTERIZATION -- TAR OIL

ımple Description:	5/11/87	6/18/87	8/10/87	9/25/87	11/16/87
ıter, wt.%	2.04	11.38*	2.30	2.69	1.41
ılorides, ppm :otal/organic)	8/4	--	4/4	2.2	1.0
emental Analysis, Dry Basis:					
ırbon, wt.%	83.83	84.80	84.50	84.76	84.19
'drogen, wt.%	8.7	9.9	8.9	8.8	8.7
trogen, ppm	8,932	5,495	6,612	6,813	8,297
ılfur, ppm	4,083	4,006	4,094	4,316	4,730
‹ygen, ppm (diff.)	61,198	43,986	55,210	52,996	57,328
:omic H/C Ratio	1.24	1.38	1.26	1.24	1.23
;C NMR, wt.%CA	61.4	--	--	--	63.8
'I Gravity	6.6	8.0	7.4	7.5	7.8
›ecific Gravity	1.025	1.014	1.019	1.018	1.016
scosity, cP @ 25C	185.1	--	--	--	--
ıur Point, F	70	--	--	--	--
h Oxide, wt. %	0.0	0.1	0.0	0.1	0.0
ltered Solids, wt.	0.25	0.60	--	0.25	.16/.20**
PSD (Microtrac), %					
<4 microns	17.6	18.3	--	12.1	7.4
4.4-6.6	17.6	12.7	--	10.3	6.8
6.6-9.4	21.2	15.6	--	13.9	9.0
9.4-13	16.8	10.3	--	16.8	10.4
13-19	17.4	12.0	--	20.3	13.4
19-27	8.4	19.1	--	17.4	11.0
>27 microns	0.8	12.0	--	8.7	41.5

Contaminated Sample
Fresh/Aged 4-Wk. @ 150F

consumption (3000-4000 SCFB). Most of the hydrogen consumed goes to saturate aromatics to meet jet fuel specifications (less than 25 percent aromatics).

Water may damage hydrotreating catalysts, and the content of hydrotreated products of materials boiling below about 300°F which may be present in jet fuels is limited. Therefore, water, along with pyridines and other heteroatomic materials boiling at less than 300°F, should be removed by distillation before hydrotreating to decrease hydrogen requirements.

Bench-scale distillations of tar oil with and without addition of toluene as an entrainer demonstrated that most of the water can be removed without use of an entrainer, as shown in Table 2.

Storage of the tar oil sample of 11/16/87 for four weeks at 150°F resulted in no significant changes in ASTM distillation and solids content, shown in Figure 5 and Table 1, respectively.

b. Phenols

The phenolic stream is characterized in Table 3. This stream is composed almost entirely of single-ring hydroxyaromatics, as shown by its high oxygen content of 16-20 weight percent. According to WRI,[1] the crude phenol can be easily hydrogenated to produce highly naphthenic JP-4 blendstock, but the hydrogen consumption is very high, around 5,000 SCFB.

Distillation results for the phenol stream, shown in Figure 6, indicate very good agreement among the samples tested. ASTM Method D-86 is used for all phenol samples. The analytical results, listed in Table 3, also show little variability with a hydrogen-to-carbon ratio of 1.21 to 1.24.

As with the tar oil, the 300°F- fraction of the phenolic stream (about 10 percent of the total) should not be hydrotreated because the water may damage catalysts, and the content of low boiling compounds is limited for jet fuel production. The phenolics stream is also stable, as shown by Figure 5.

TABLE 2

GREAT PLAINS COAL LIQUIDS UPGRADING

TAR OIL BENCH-SCALE DISTILLATIONS RESULTS

METHOD	WATER, WT. %
KARL FISHER ANALYSIS, AS-RECEIVED TAR OIL	1.41
AZEOTROPIC DISTILLATION, TOLUENE	1.28
VACUUM DISTILLATION, NO ENTRAINER	1.21

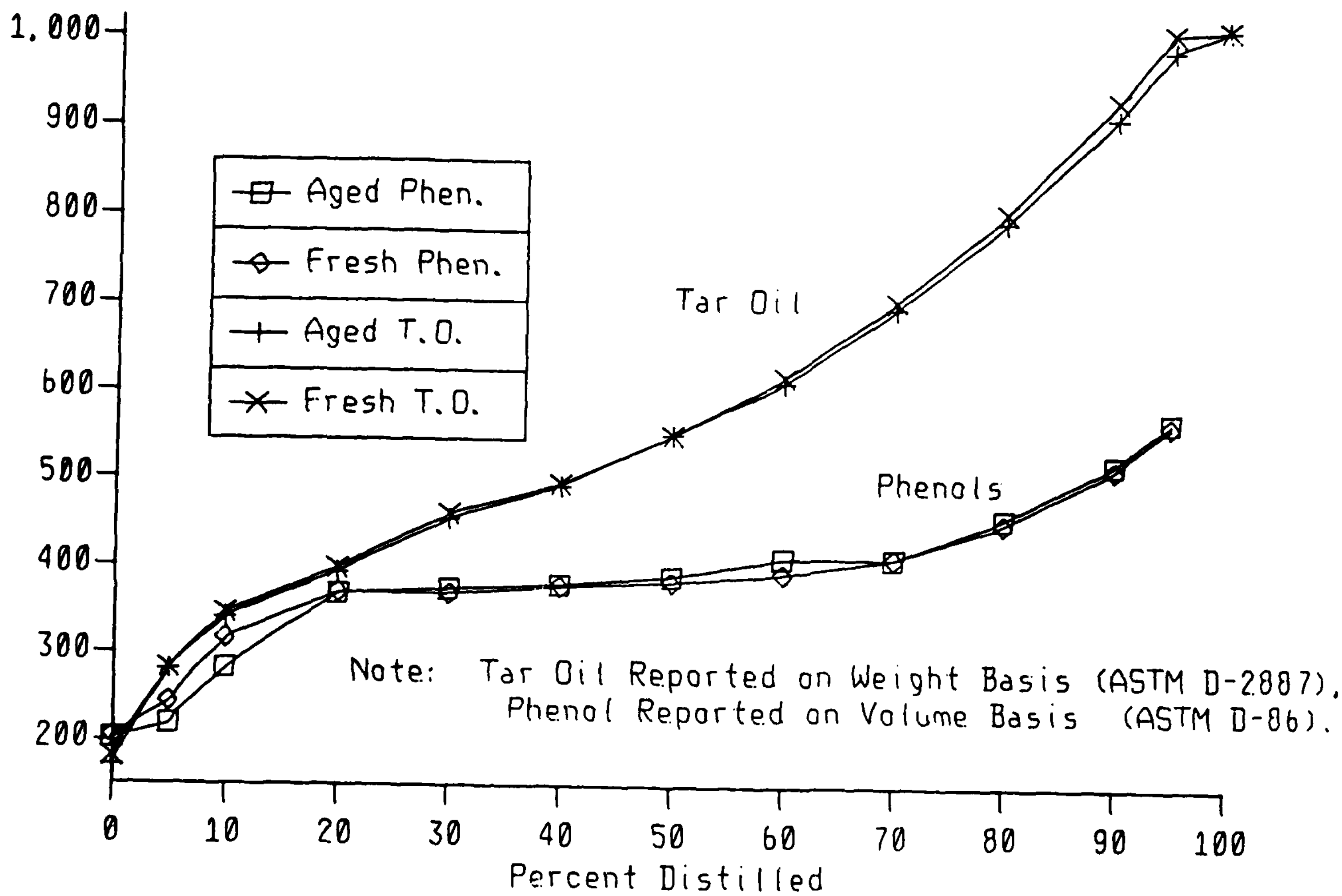

FIGURE 5
GREAT PLAINS COAL LIQUIDS UPGRADING
SIMULATED DISTILLATION RESULTS
BEFORE/AFTER 4-WEEKS STORAGE @ 150 F

TABLE 3

GREAT PLAINS COAL LIQUIDS UPGRADING

FEEDSTOCK CHARACTERIZATION -- PHENOLS

mple Description:	5/11/87	6/18/87	8/10/87	9/25/87	11/16/87
ter, wt.%	--	--	5.46	5.86	5.56
lorides, ppm otal/organic)	--	4/5	5/3	6.8	<1
emental Analysis, Dry Basis:					
rbon, wt.%	72.88	72.38	76.76	76.76	75.89
drogen, wt.%	7.6	7.4	7.8	7.8	7.7
trogen, ppm	5,360	5,060	4,876	4,748	4,892
lfur, ppm	780	570	751	914	837
ygen, ppm (diff.)	189,060	197,070	148,593	148,789	158,503
omic H/C Ratio	1.24	1.21	1.21	1.21	1.21
C NMR, wt.%CA	82.6	--	--	--	--
I Gravity	1.6	2.3	1.4	0.9	0.1
ecific Gravity	1.063	1.058	1.065	1.069	1.075
scosity, cP @ 25C	15.3	--	--	--	--
ur Point, F	-40	--	--	--	--

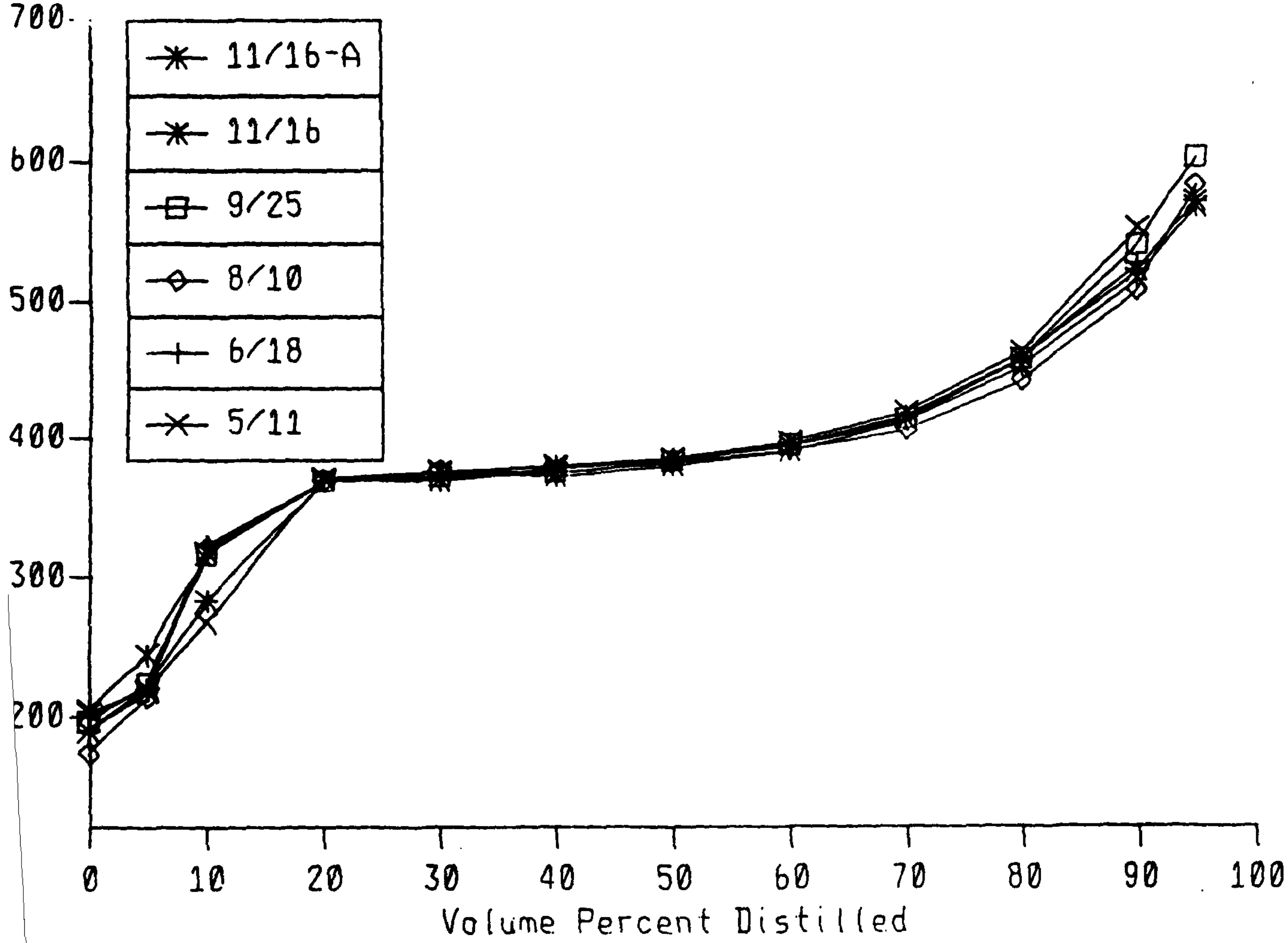

FIGURE 6
GREAT PLAINS COAL LIQUIDS UPGRADING
PHENOLS DISTILLATION RESULTS
(Method: ASTM D-86)

c. Naphtha

The naphtha stream is characterized by analyses shown in Table 4 and distillation results in Figure 7. Comparison of as-received and caustic-extracted samples shows significant decreases in carbon-to-hydrogen ratio, sulfur, and nitrogen for extracted naphtha. As-received samples contain significant amounts of methanol, acetone, and methylethylketone from the Great Plains Rectisol unit, and the presence of thiols, mercaptans, and thiophenes. These produce extremely noxious odors when opened. Resulting safety and environmental concerns have limited the number of naphtha analyses and distillations performed. The boiling range of the naphtha stream is too low to produce significant amounts of jet fuel. Aromatics can be recovered from this stream, but only after hydrotreatment to reduce sulfur and nitrogen levels.

2. Subtask 1.2 Process Modelling and Design

The seven design cases for which conceptual process designs have been developed are listed in Figure 2. The first six cases, which are required by the contract, involve making jet fuel either at maximum production rate or with co-products such that the total product slate is profitable. The seventh case has been added by Amoco as a natural extension of the project. This case drops the requirement of production of jet fuel and produces the most profitable product slate.

TABLE 4

GREAT PLAINS COAL LIQUIDS UPGRADING

FEEDSTOCK CHARACTERIZATION -- NAPHTHA

Sample Description:	5/11/87	5/11/87 Extract	6/18/87	6/18/87 Extract	8/10/87
Water, wt%	0.56	--	0.42	0.11	0.51
Elemental Analysis, Dry Basis:					
Carbon, wt%	84.22	86.33	84.46	86.30	84.85
Hydrogen, wt%	10.00	9.57	9.97	9.93	9.94
Nitrogen, ppm	2,062	630	2,089	1,321	1,970
Sulfur, ppm	17,397	12,600	16,871	861	18,394
Oxygen, ppm (diff)	38,365	27,770	36,674	35,459	31,702
Atomic H/C Ratio	1.41	1.32	1.41	1.37	1.40
13C NMR, wt% CA	65.9	--	--	--	--
API Gravity	38.9	39.5 40.0	39.8	53.8	
Specific Gravity	0.830	0.827	0.825	0.826	0.764
Viscosity, cP @ 25C	0.5	--	--	--	--
PONA Analysis, Liq. Vol%					
Paraffins	23.1	7.4	8.3	--	--
Naphthenes	12.2	3.9	4.5	--	--
Olefins, noncyc/cy	--	14.4/9.7	15.6/7.7	--	--
Aromatics	64.6	64.6	63.9	--	--

TABLE 4 (concluded)

GREAT PLAINS COAL LIQUIDS UPGRADING

FEEDSTOCK CHARACTERIZATION -- NAPHTHA

Sample Description:	5/11/87		5/11/87 Extract	6/18/87	6/18/87 Extract	8/10/87	
Light Ends, wt%		(bomb)					
C1	--	--		--	--	0.0	
C2	--	--		--	--	0.0	
nC3	0.0	0.1		0.0	--	0.0	
C3=	--	--		--	--	--	
nC4	0.1	0.2		0.1	--	0.1	
iC4	0.0	0.0		0.0	--	0.0	
C4=	0.438	1.064			0.725	--	0.886
nC5	0.469	0.603			0.217	--	0.259
iC5	0.158	0.196			0.175		
C5 cyc.	0.241	0.280		0.253	--	0.291	
C5=	2.258	2.735		2.046	--	3.003	
C5==	3.507	4.099		2.887	--	2.932	
Misc. Components, wt%							
Methanol	2.27		--	2.47	--	1.82	
Acetone	6.32		--	--	--	--	
MEK	3.32		--	--	--	--	
Benzene	45.62		--	46.04	--	46.30	
Toluene	17.93		--	15.99	--	16.54	
Xylenes (incl. EB)	3.25		--	3.60	--	3.86	
C5==	3.507		4.099	2.887	--	2.932	
Misc. Components, wt%							
Methanol	2.27		--	2.47	--	1.82	
Acetone	6.32		--	--	--	--	
MEK	3.32		--	--	--	--	
Benzene	45.62		--	46.04	--	46.30	
Toluene	17.93		--	15.99	--	16.54	
Xylenes (incl. EB)	3.25		--	3.60	--	3.86	

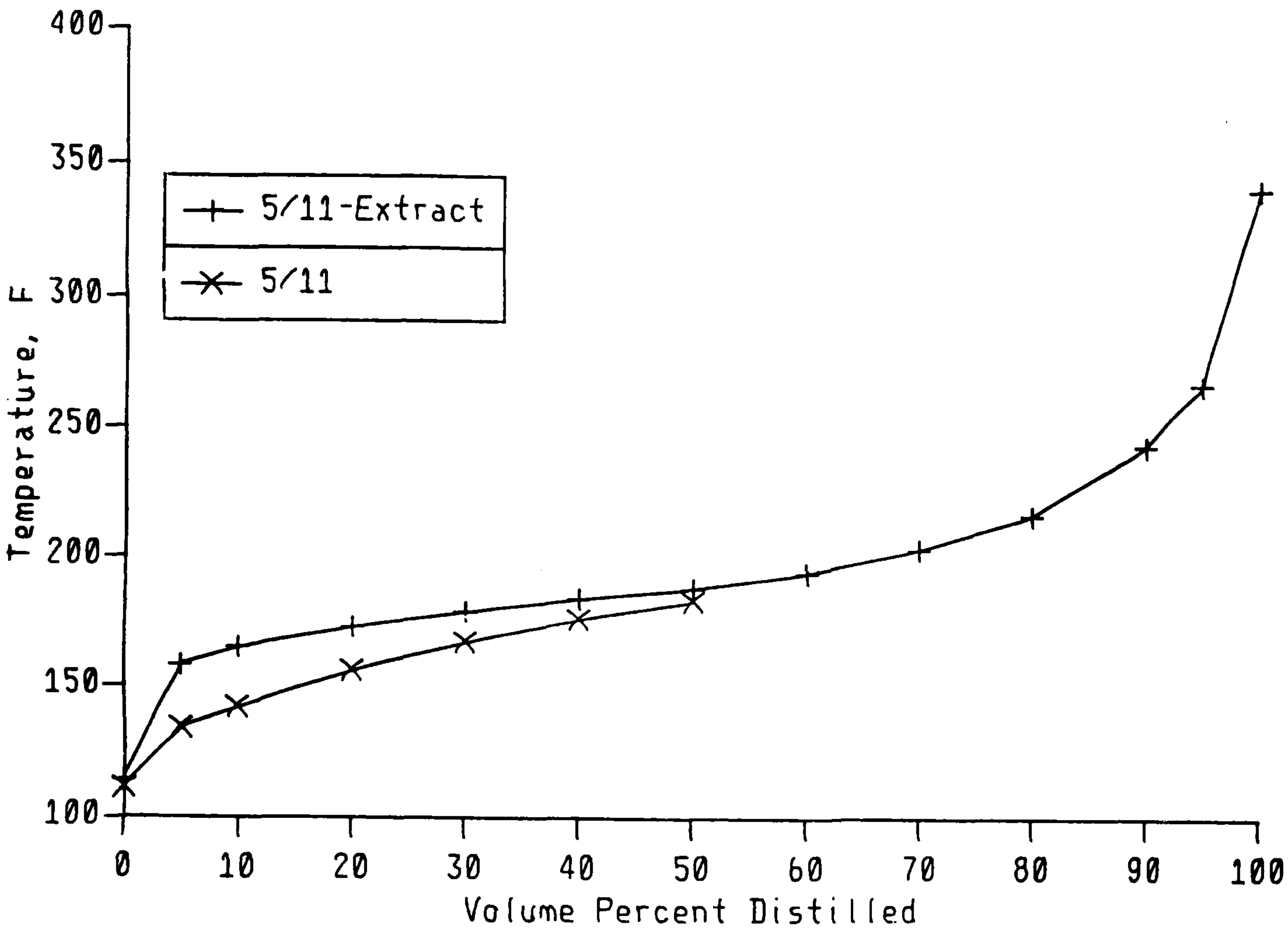

FIGURE 7
GREAT PLAINS COAL LIQUIDS UPGRADING
NAPHTHA DISTILLATION RESULTS
(Method: ASTM D-86)

Many alternative routes to upgrade the by-product streams can be proposed. While it may be easy to pick the route that maximizes jet fuel production, the routes that produce a profitable product slate with jet fuel, or the maximum profit product slate are difficult to determine without using a computerized approach which can consider many alternatives for the profitable cases.

a. Technical Approach

To evaluate all possible cases, Amoco has developed a linear program for the Great Plains by-products. This LP has numerous inputs, including:

1. Amoco's proprietary linear programming technology (including refinery process models).

2. The results of the Western Research Institute scoping hydrotreating study.[1]

3. The price structure for various oxygenated chemicals from the Sinor report.[2]

4. Amoco's price structure for hydrocarbon products and feeds in the North Dakota area.

5. Information on processes to upgrade oxygenated chemicals.[4,5]

The Great Plains LP contains 1,450 equations and approximately 2,400 variables. The equations describe process and utility constraints, mass and energy balance constraints, blending options, capital and operating costs.

Linear programming is used by Amoco to help guide research and refining planning. It is a methodology to rigorously analyze the effects of process

technology, feeds, and products on refinery configurations. Amoco's linear program technology is based on Amoco's extensive commercial petroleum refining experience. It can be used to evaluate additions and revamps of existing refineries or design a new refinery (utilities available), as would be the case for Great Plains.

Amoco's linear programming technology has an extensive data base. It includes process models; raw material and product specifications, and prices; capital investment, operating cost and utility requirement correlations; and product blending algorithms. The process models are based on Amoco's technology wherever possible, but all technologies are commercially proven and licensable. The data base is used in conjunction with a refinery and petrochemical modeling system that draws on the data base to construct linear program models, which are solved consistent with whatever constraints the user chooses to apply. In essence, a tailor-made LP model is developed for each new problem. The user may specify virtually any combination of feed material and/or product volumes, specifications, and prices. Alternatively, or in addition, refinery configurations can be specified completely, in part, or not at all. Optimal solutions are developed for each variable not specified. This affords a great deal of flexibility in analyzing a wide variety of refining environments and refinery configurations. The LP analysis also considers and optimizes process interactions and product blending effects.

Table 5 lists the processes which are included in the linear program developed for Great Plains liquid by-product upgrading. The major process blocks are also shown schematically in Figure 8. Several processes use standard petroleum refining technology. Most other processes (denoted by (1) in Table 5) have been adapted from standard petroleum technology by adjusting for the different feedstock inspections of the Great Plains streams. Where interpolation between literature or proprietary data points is not possible, proprietary Amoco process simulation models are used to predict the effects of feedstock inspections on process operations. These simulations have been developed from data using petroleum based feeds, including high-aromatics stocks similar to the Great Plains streams. A few processes (denoted by (2) in Table 5) are critical enough or different enough from standard petroleum

processing that customized models were developed. The development of these models is outlined below.

TABLE 5

PROCESS BLOCKS INCLUDED IN LINEAR PROGRAM SIMULATION OF GREAT PLAINS LIQUIDS UPGRADING

Aromatics Recovery (1)
Butane and Pentane Isomerization
Catalytic Cracking (1)
Cresylic Acid Fractionation (1)
Delayed Coker (1)
Distillate Hydrotreater (NiMo & NiW) (2)
Dynaphen (2)
Gas Oil Desulfurizer
Gasoline Reformer (1)
Hydrocracking for JP-4, JP-8, and Gasoline (1)
Hydrogenation/Saturation (2)
Naphtha Distillation (1)
Naphtha Hydrotreater (2)
Naphtha Sweetening
Phenoraffin/ANG Extraction (2)
Pressure Swing Absorption (PSA) (1)
Product Blending
Propylene, Butylene, and Amylene Alkylation
Propylene Concentration
Propylene and Butylene Polymerization
Sulfur Recovery
Tar Oil Distillation (1)
Utilities Generation

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(1) Units which differ in feedstock properties from standard petroleum refining technology, but can readily be adapted to Great Plains stocks.

(2) Units which differ enough from standard petroleum refining technology to be customized for application to Great Plains stocks (see text for details).

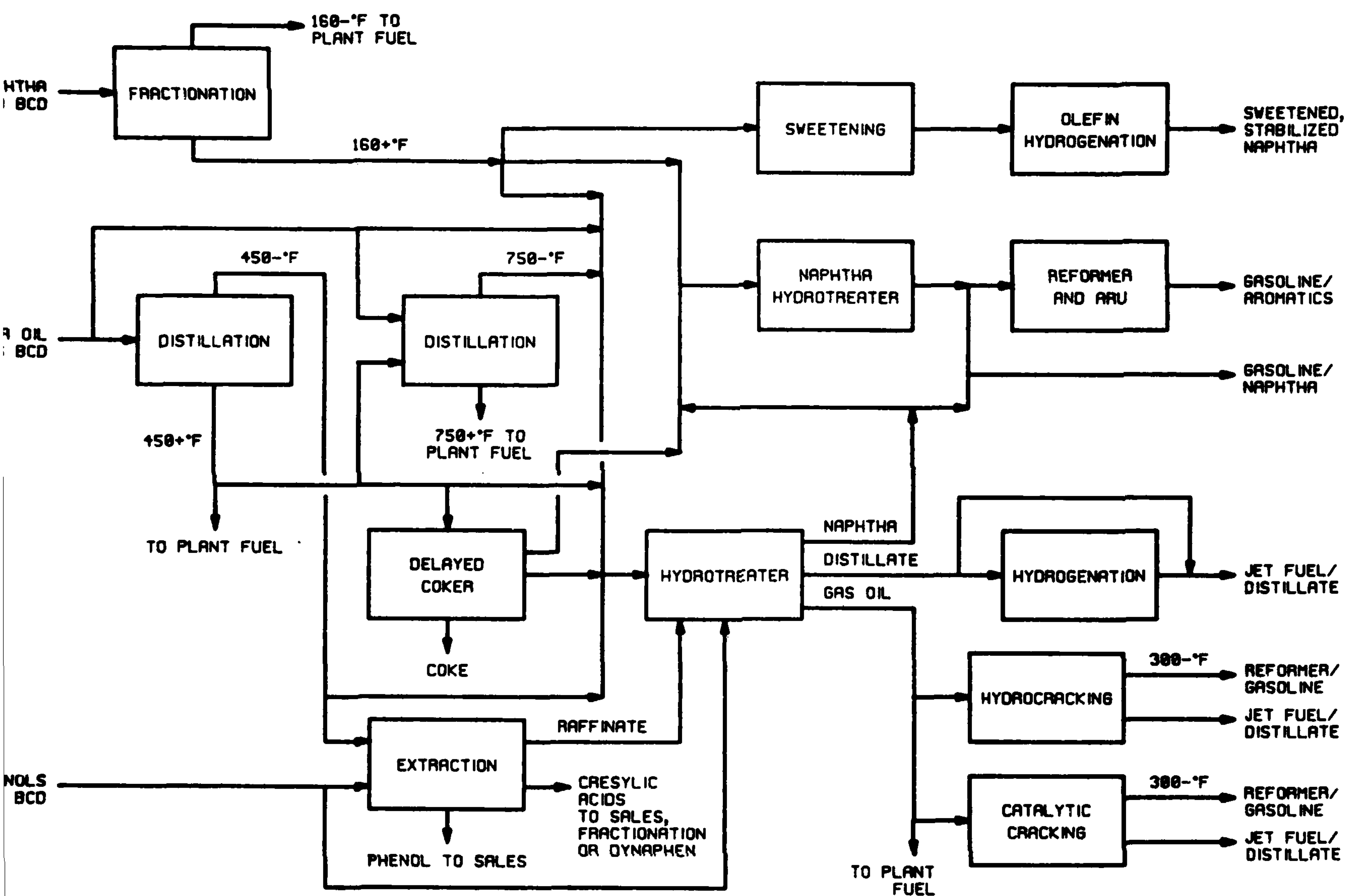

FIGURE 8
LINEAR PROGRAM FLOWSHEET FOR GREAT PLAINS BY-PRODUCT REFINING

The distillate hydrotreating and hydrogenation/saturation processes are modeled using process data from coal-derived feedstocks published by Chevron.[6] Since aromatics saturation reactions are so important in processing Great Plains streams to meet jet fuels specifications, an aromatics saturation model was developed and reported in Appendix A. The phenol hydrotreating is based on WRI[1] test 87-07-5, adjusted for different feed inspections and adding light ends yields. The hydrotreating results from neutral oils from Phenoraffin extraction are estimated based on a higher aromatics content than the original streams.

An expanded bed technology was selected to hydrotreat the tar oil. This process selection was based on the following advantages for expanded beds, relative to fixed beds, for this feed: (1) Little or no feed preheat is required, (2) Relatively isothermal operation with no large reactor temperature exotherm and relatively simple reactor temperature control requirements, (3) Tolerance to solids in the feed, and (4) Operating flexibility.

The data used for estimating the yields and process conditions for hydrotreating the Great Plains naphtha comes from Chevron's results with SRC-II naphtha.[6] A proprietary fractionation simulation is used to divide the Great Plains naphtha stream into 160°F- and 160°F+ cuts. The 160°F- cut is used as fuel. The coker naphtha hydrotreating yields are based on proprietary data using delayed coker naphtha from petroleum stocks.

All hydrotreater (naphtha and tar oil) investments and operating costs are based on estimates developed by Lummus, based on their preliminary process designs. Dynaphen processing is based on the literature.[5] The Dynaphen process, which is still to be confirmed for Great Plains feedstocks by HRI, will partially dealkylate coal liquids to phenol and benzene.

The Phenoraffin process yields are based on published recoveries of pure components adjusted for Great Plains phenolic composition. Phenoraffin is Lurgi's proprietary process for separation and purification of phenols, cresols, and other commodity chemicals from mixed oxygenate streams. A firmer process design and cost basis for Phenoraffin was sought from Lurgi, but they

reportedly needed to do experimental work using Great Plains feedstocks before they would provide this information. As a result, Lummus developed a process design and cost basis for the phenolic extraction based on a dual solvent extraction process under development by ANG.[4] Henceforth in this report, this process will be referred to as the Phenoraffin/ANG process.

Feed costs and product values are taken from Amoco proprietary North Dakota refinery information and a recent Sinor study.[2] Hydrogen cost, based on extraction from Great Plains syngas, is from a recent Burns and Roe study.[8] Hydrogen is extracted by pressure swing adsorption. The cost of feeds and the value of products used in the linear program are listed in Table 6.

Product specifications for the various grades of gasoline and jet fuel have also been included in the LP data base. These specs are given in Table 7. Specification for the oxygenated chemical products have not been determined for the preliminary evaluation in Task 1. It is assumed that the Phenoraffin/ANG process which produced the various oxygenated chemicals, could meet the required specifications. This assumption should be verified.

b. Conceptual Upgrading Schemes

Using the Great Plains LP, conceptual upgrading schemes have been developed, along with sensitivities to key variables and assumptions. The cases are presented in the order: maximum jet fuels (Case 1:JP-4; Case 3:JP-8; Case 5:JP-8X); maximum profit (Case 7); and profitable jet fuels (Case 2:JP-4; Case 4:JP-8; Case 6:JP-8X).

The maximum jet fuels cases have been estimated by hand-calculation and by the LP. The hand calculation allowed Lummus to begin work before the LP was finished. They also provide a check on the accuracy of the LP. The economics presented below, however, are based on the LP estimated cases. The LP simulations do not exactly match the hand-calculated process schemes because of three reasons. First, the linear program cannot exactly match the non-linear curves which relate yields and qualities to process conditions, because it approximates these curves with straight line segments. Second, the

TABLE 6

COST AND PRICE BASES AND ASSUMPTIONS

Input Streams	Price $/Bbl	Max. BCD	Fuel Value, $/MMBtu
Natural gas	13.57	-	2.15
LPG/Propane	7.57	-	2.15
i-Butane	19.11	-	4.98
n-Butane	11.76	-	2.95
GP Naphtha	12.69	660	2.15
GP Phenols	10.43	833	2.15
GP Tar Oil	13.07	2896	2.15
Toluene	38.01	5000	
GP Syngas for H2	1.23/MSCF H2	36810 MSCFD	2.47

Output Streams			Limitation:
LPG/Propane	7.56	-	
Unleaded Gasoline	23.35	-	
Unleaded Premium	26.29	-	
Sweetened GP Naphtha	25.45	-	
Reformer Feed	24.61	-	
Hydrotreated GP Naphtha	30.00	-	
JP-4	24.19	-	
JP-8	21.84	-	
JP-8X	21.84	-	
Benzene	48.00	-	
Toluene	38.00	-	
Xylene	49.00	-	
Phenol	80.00	-	
o-Cresol	182.00	25	10% U.S. Market
m,p-Cresol	199.00	80	10% U.S. Market
Xylenols	171.00	30	10% U.S. Market
Cresylic Acids	134.00	140	10% U.S. Market
GP Fuel Pool	2.15/MMBTU	25540 MMBTU/CD	
Sulfur	125/LT	-	
Coke	16/T	-	

TABLE 7

PRODUCT SPECIFICATIONS

Jet Fuel Specification	JP-4	JP-8	JP-8X
Boiling Range, °F	160-518	300-572	300-572
Specific Gravity, min.	0.7507	0.775	0.850
" " , max.	0.8017	0.840	
Sulfur, wt%, max.	0.4	0.3	0.3
Aromatics, min.			10
" , max.	25	25	30
Paraffins, max.			10
Naphthenes, min.			70
" , max.			90
Reid Vapor Pressure,min.	2.0		
" " " ,max.	3.0		
Flash Point, °F, min.		100	122
Pour Point, °F, max.		-72	-62

Gasoline Specification	Unleaded Regular	Unleaded Premium	Unleaded Midgrade
Road Octane, min.	87	93	89
Reid Vapor Pressure,min.	9.6	9.6	9.6
" " " ,max.	12.7	12.7	12.7
Specific Gravity, max.	0.7669	0.7669	0.7669
Sulfur, wt%, max.	0.1	0.1	0.15

In addition, distillation specifications consistent with ASTM D-439 were employed.

In addition, distillation specifications consistent with ASTM D-1655, MIL-T-5624L, and MIL-T-83133 were employed. Pour point and flash point specifications were met for JP-8 and JP-8X using proprietary non-linear blending techniques.

Also sold but blended from stocks which are produced to required purity specifications are: LPG, sweetened naphtha, reformer feed, hydrotreated naphtha, BTX, phenol, o-cresols, m- and p-cresols, xylenols, cresylic acids, coke, and sulfur. See Table 3.

LP adjusts process conditions in response to optimal costs until process limitations are reached. Thus, the LP design always results in a solution which is limited by some constraint. Finally, a true LP maximum sometimes results in absurd processing choices. For example, the unconstrained LP maximum JP-4 case includes processing the Great Plains naphtha and several other streams, of less than 100 B/D, to get a few more barrels of jet fuel. It does this by building a very small catalytic cracker to get a low gravity jet fuel stock which allows a few more barrels of high gravity stocks to go to jet fuel. The cases reported here externally forbid process options which result in very high costs or very small stream or unit sizes. The prohibited streams are less than 10 percent of the parent stream volumes.

The other four cases have no corresponding hand-calculated solutions to compare with, since they are economic optima. Some external process limits are also imposed on these cases to avoid processing very small streams or building very small units.

Lummus's estimate of investment and operating costs for each case are reported below.

Case 1 - Maximum JP-4.--Figure 9 shows hand-calculated maximum JP-4 case. This case hydrotreats the Great Plains tar oil and phenol streams, while leaving the naphtha for plant fuel needs. The hydrotreated 500°F+ stream is hydrocracked to produce additional jet fuel blending stocks. Combined JP-4 production is 4270 BSD, along with 560 BSD of naphtha stocks. The LP solution, shown in Figure 10, produces about 4 percent more JP-4 than the hand-calculated case. This is within the range expected, due to LP limitations discussed above. Figure 9 reports flows in BSD (barrels per stream day), while Figure 10 reports them in BCD (barrels per calendar day). Using the Great Plains operating factor of 91 percent, BCD values are 9 percent lower than BSD values. Yields of key products, in barrels per calendar day, are compared in Table 8. Both Figures 9 and 10 include the same processes, so the match of these two cases is sufficient for this scoping evaluation.

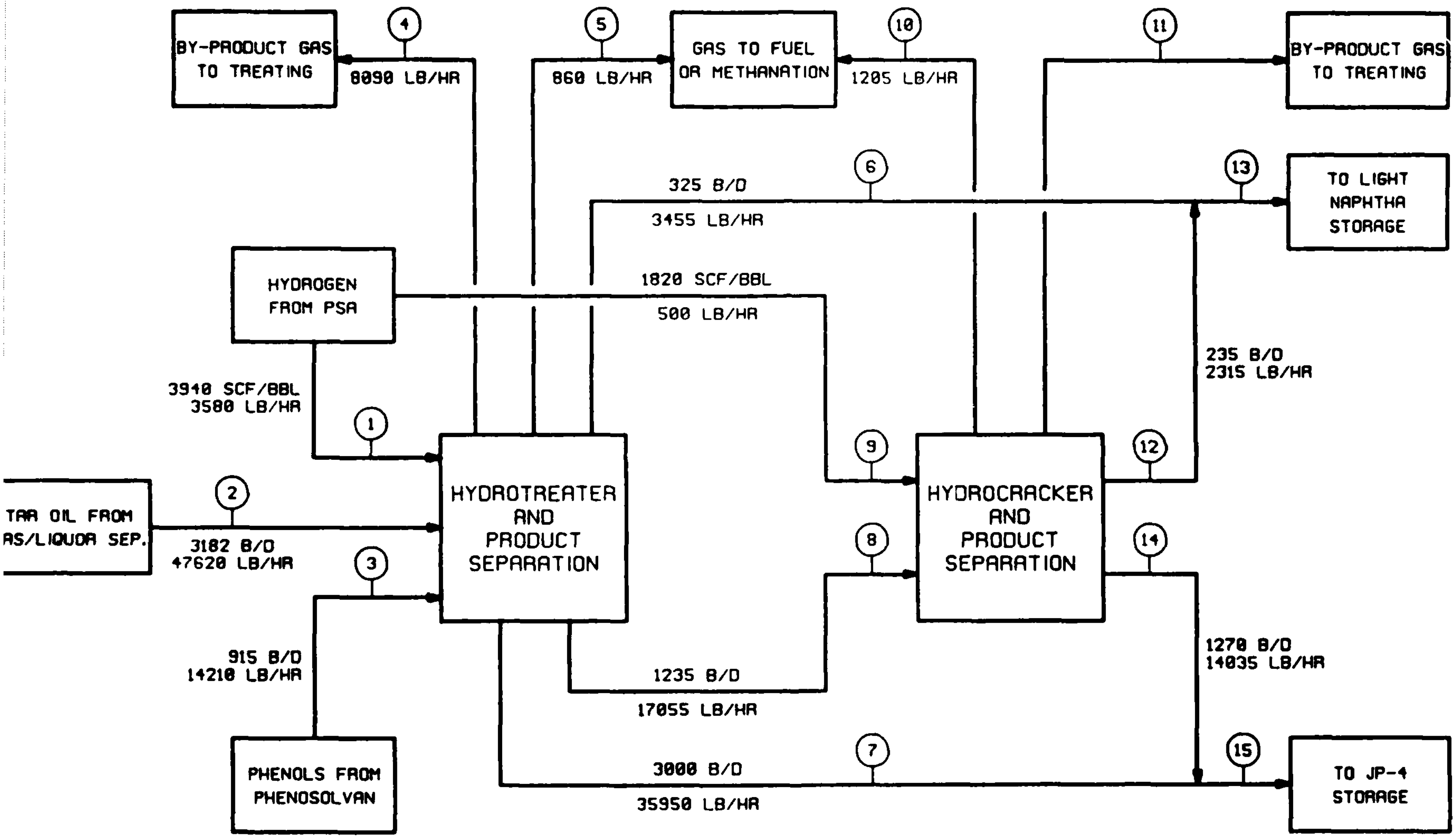

FIGURE 9
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 1: MAXIMUM JP-4 PRODUCTION

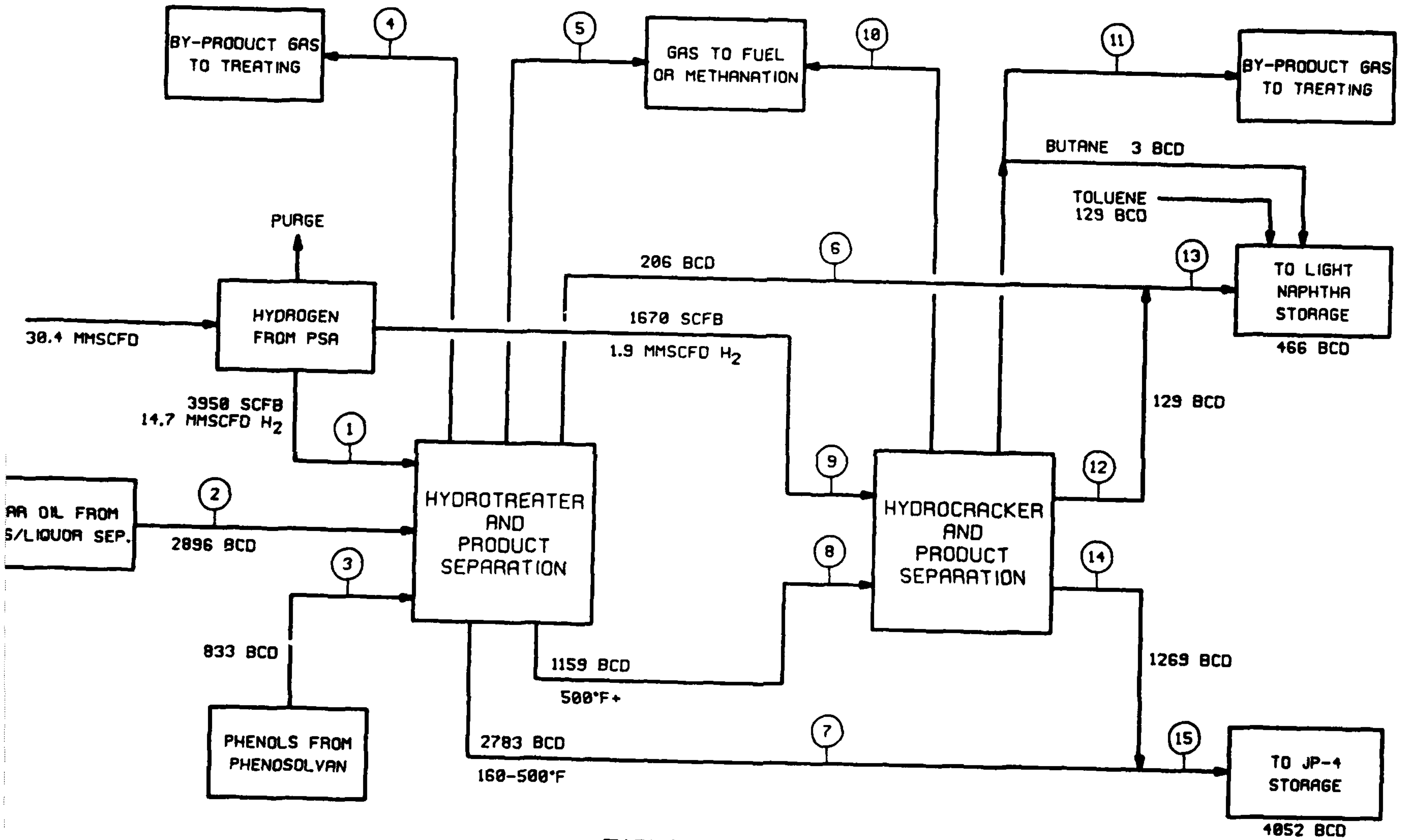

FIGURE 10
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 1: LP MAXIMUM JP-4 PRODUCTION

TABLE 8

COMPARISON OF MAXIMUM JET FUEL CASES WITH LP SOLUTIONS

Streams	JP-4 Hand-Calculated	JP-4 LP	JP-8 Hand-Calculated	JP-8 LP	JP-8X Hand-Calculated	JP-8X LP
Naphtha, BCD	510	466	1215	1222	505	378
Jet Fuel, BCD	3886	4052	2266	2323	1784	1764
Fuel Oil, BCD	0	0	0	0	804	947
Syngas, MMSCFCD	31.0	30.4	24.0	23.7	13.6	14.3

Case 3 - Maximum JP-8.--Figure 11 shows hand-calculated maximum JP-8 case. This case uses only the tar oil stream to make jet fuel, while leaving both the phenol and naphtha stream for plant fuel needs. The same sequence of hydrotreating and hydrocracking is used as for JP-4 production, but the cut point from the hydrotreater is 550°F instead of 500°F. The linear program maximum JP-8 case, shown in Figure 12, again estimates about 3% more JP-8 than the hand calculated case, as Table 8 shows. Again, the close match of unit operations in Figures 11 and 12 confirms the estimating design basis.

Case 5 - Maximum JP-8X.--Figure 13 shows the hand-calculated maximum JP-8X case. Here too, only the Great Plains tar oil stream is used for jet fuel production. This case first fractionates the tar oil at 750°F and hydrotreats only the 750°F- cut. The hydrotreater products are naphtha, JP-8X, and a 550°F+ stock which is combined with 750°F+ tar oil to provide plant fuel. The high-boiling fractions of the tar oil are too aromatic to make JP-8X. The LP maximum JP-8X case, shown in Figure 14, bypasses some tar oil directly to hydrotreating to save distillation costs. However, the close match of units chosen and overall yields listed in Table 8 again confirms the design basis used. The preferred design basis should be to fractionate all the tar oil, since the incremental cost of this capacity is low and this improves plant operating flexibility.

Case 7 - Maximum Profit.--The maximum profit case is shown in Figure 15. This case processes only the Great Plains naphtha and phenol streams. The tar oil stream is used as plant fuel. The naphtha is fractionated to remove undesirable heteroatoms boiling below 160°F. Improved operation of the existing Great Plains stripper might eliminate this process step, but it is included here if the desired stripper performance cannot be achieved. The 160°F+ naphtha is hydrotreated and then aromatic components (benzene, toluene, and xylene, or BTX) are recovered for sale. The remaining naphtha components are blended to gasoline. A small PSA unit recovers 0.2 MMSCFD of hydrogen from the Great Plains Rectisol gas stream to provide the hydrogen for this hydrotreating step.

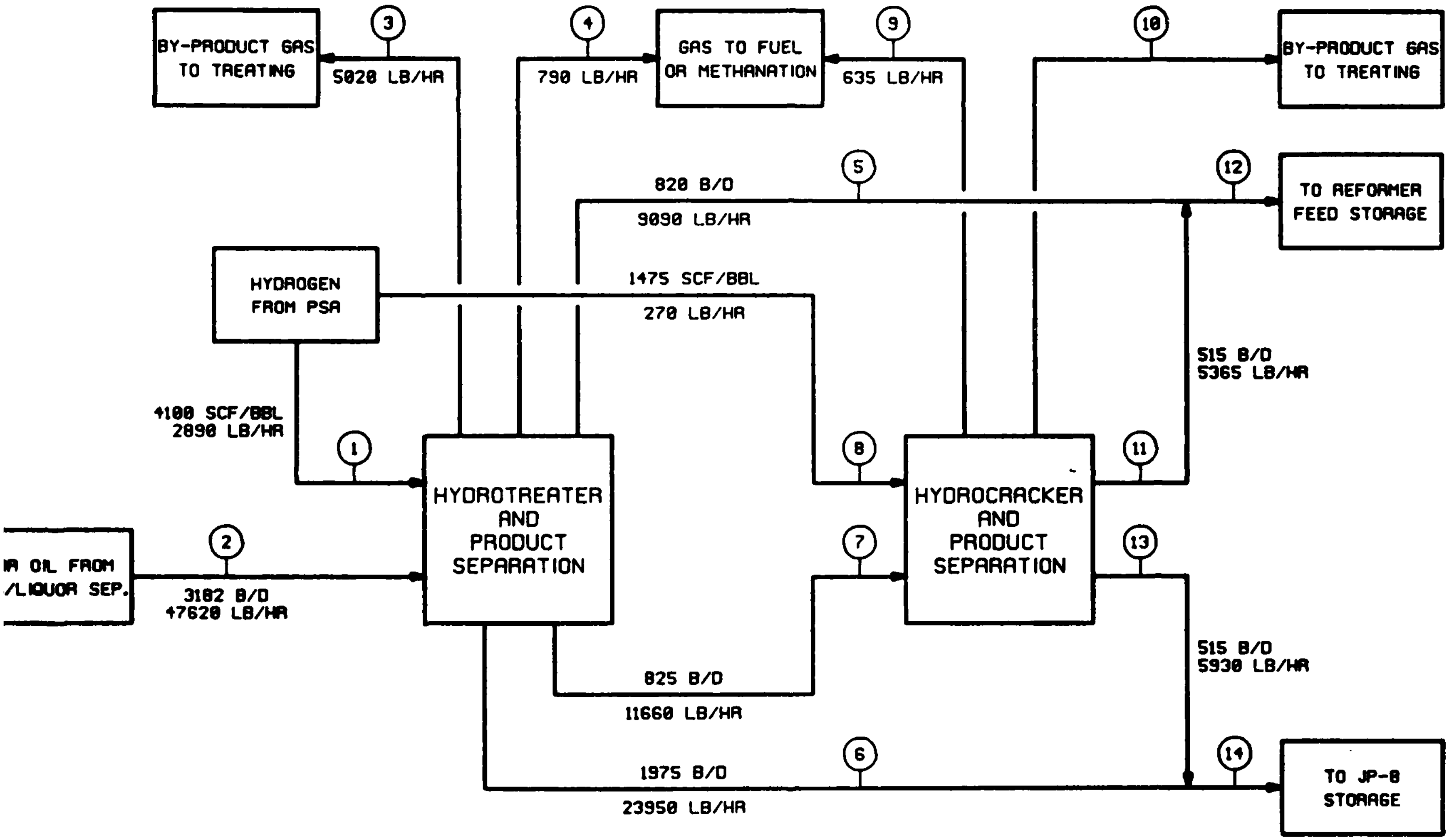

FIGURE 11
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 3: MAXIMUM JP-8 PRODUCTION

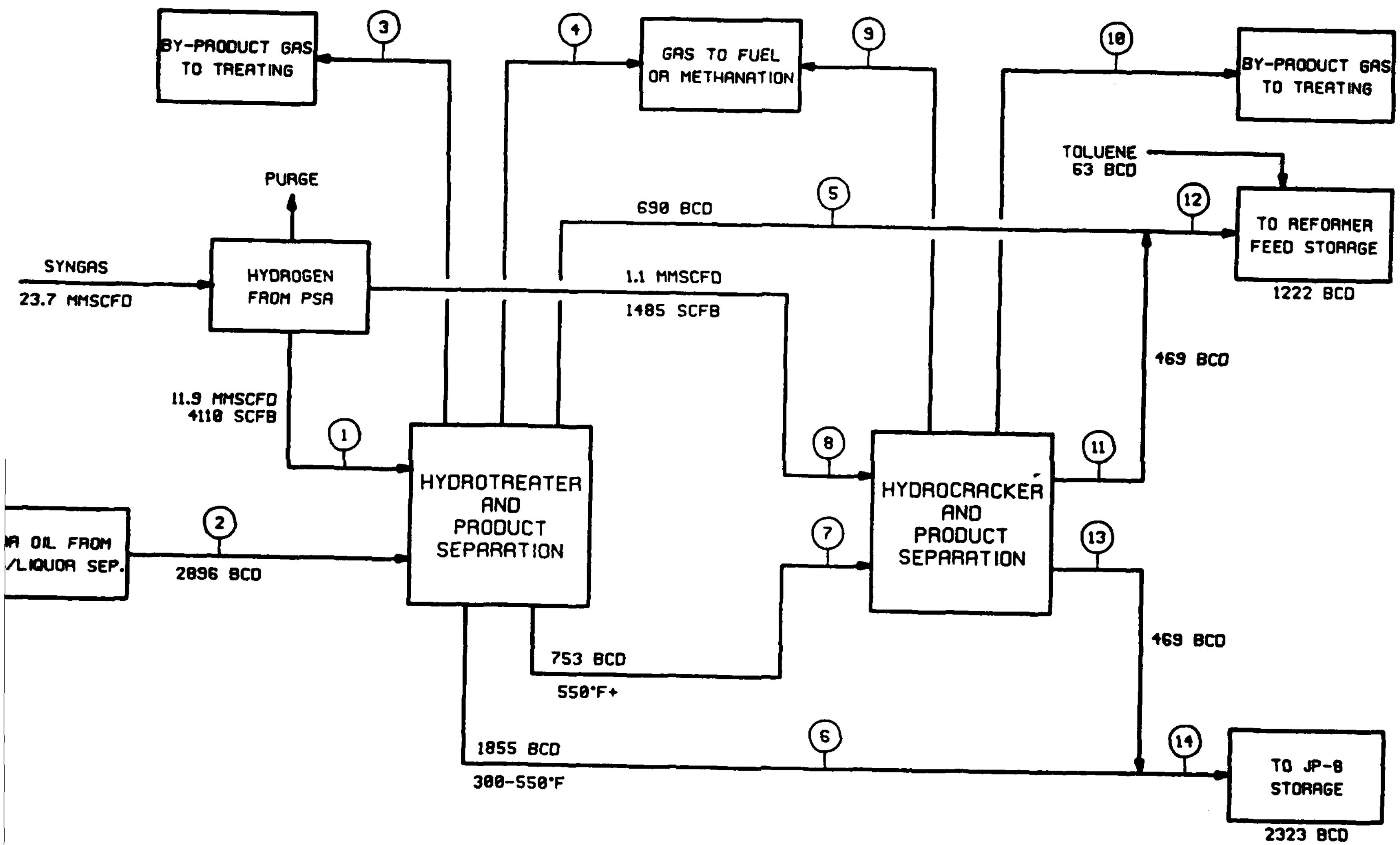

FIGURE 12
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 3: LP MAXIMUM JP-8 PRODUCTION

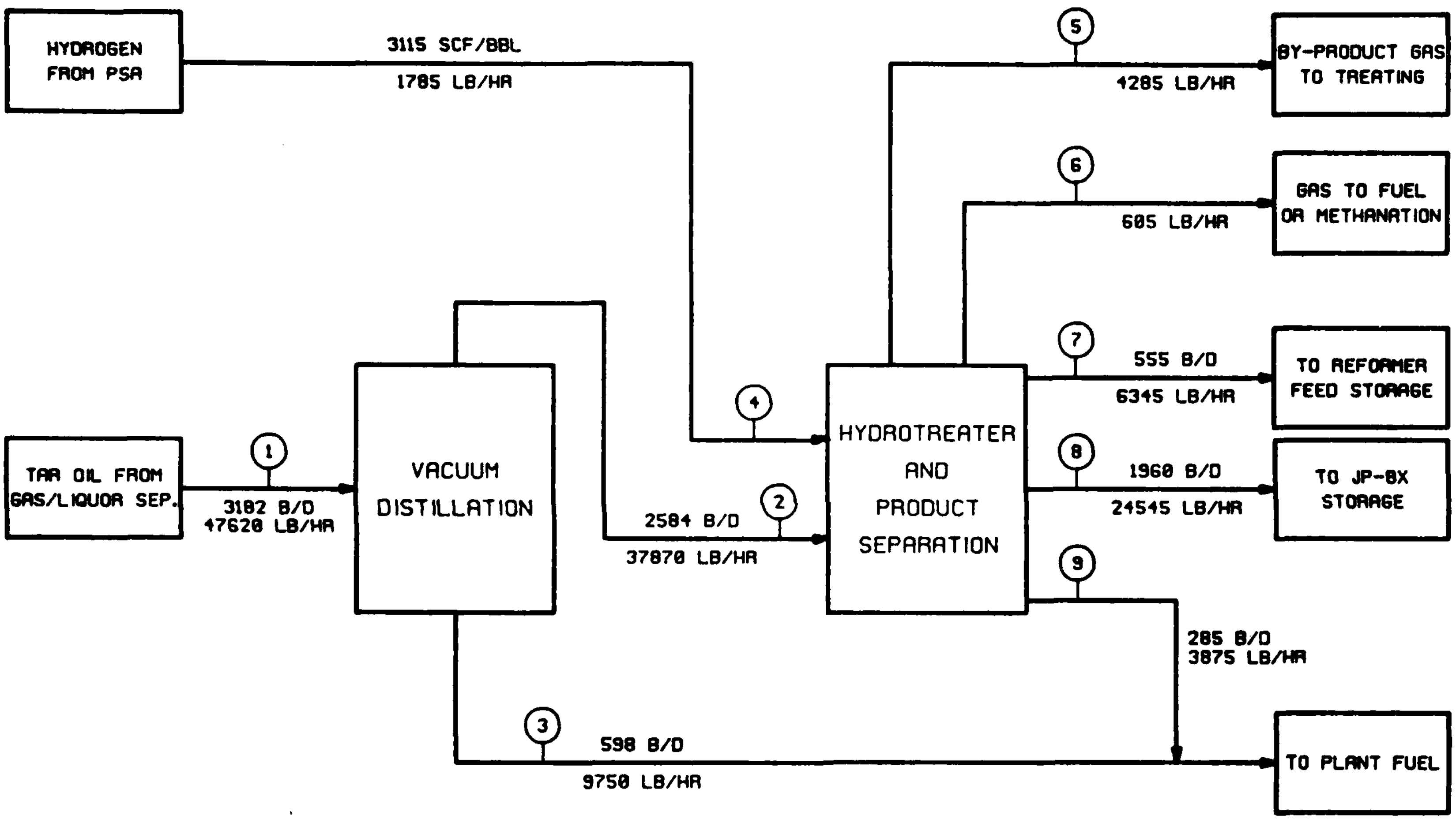

FIGURE 13
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 5: MAXIMUM JP-8X PRODUCTION

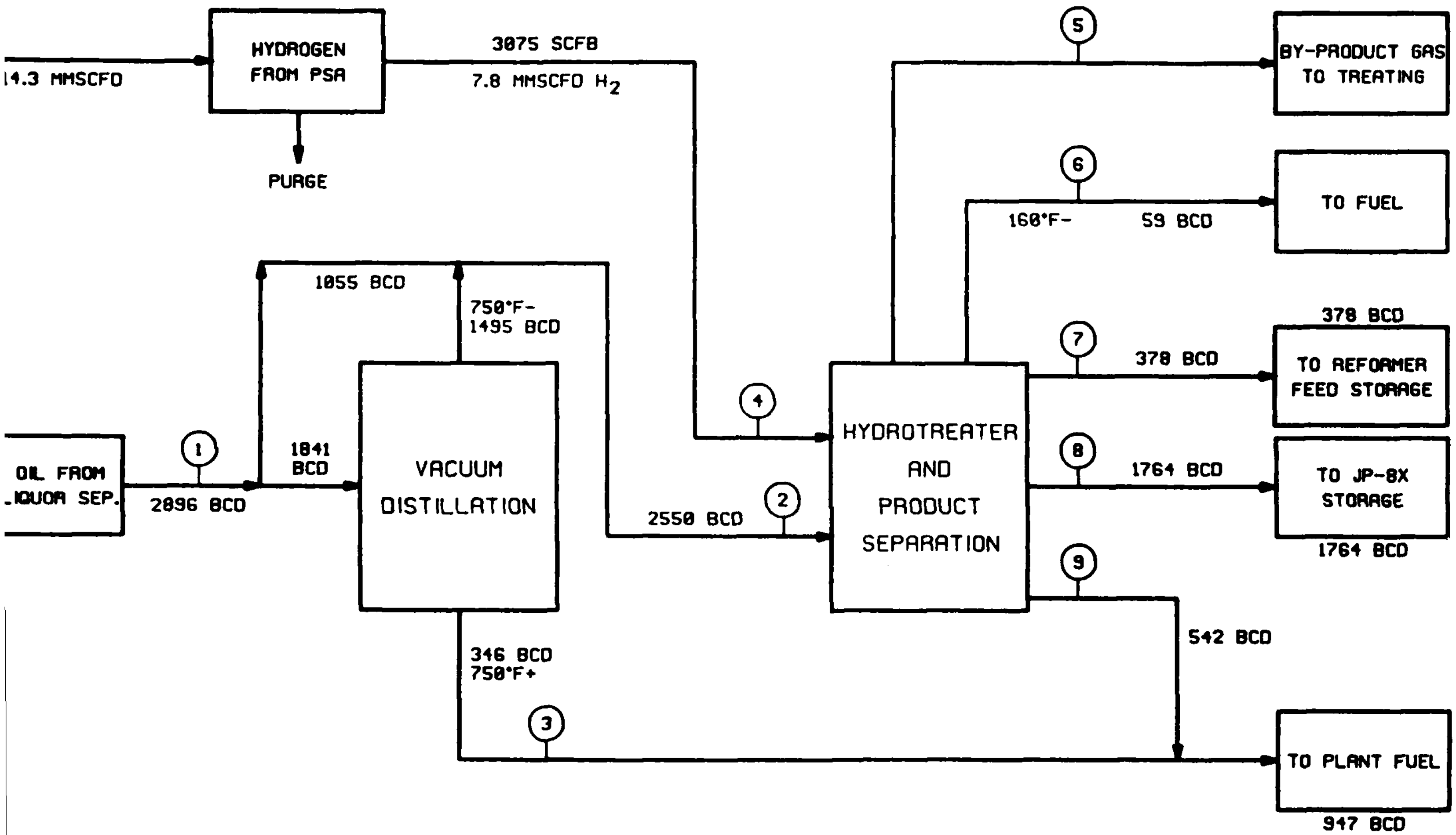

FIGURE 14
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 5: LP MAXIMUM JP-8X PRODUCTION

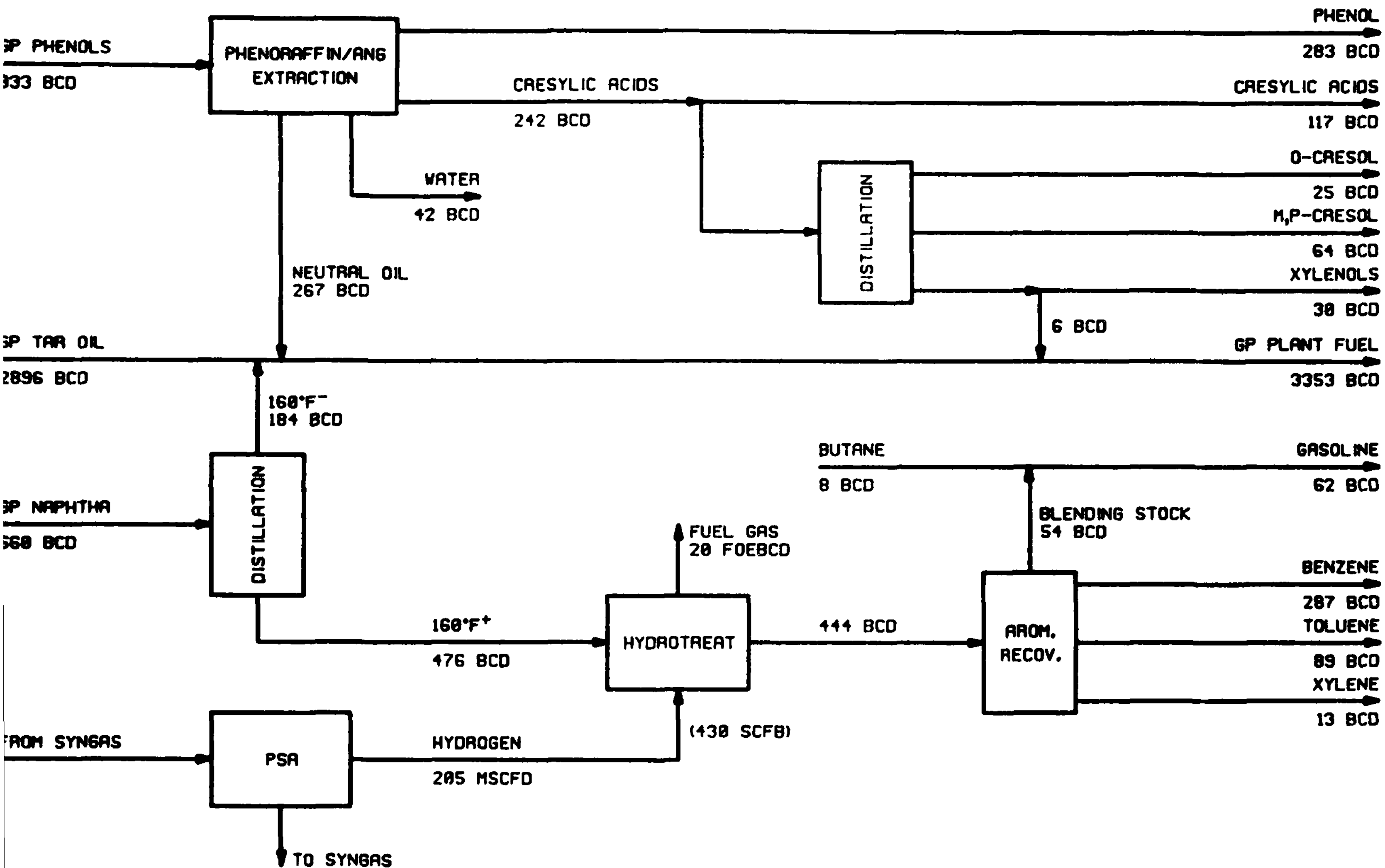

FIGURE 15
GREAT PLAINS LIQUID BY-PRODUCTS USE
CASE 7: MAXIMUM PROFIT CASE

The Great Plains phenol stream goes to a Phenoraffin/ANG extraction unit to recover purified phenol and cresylic acids. The remaining neutral oil stream is burned for plant fuel. Some of the cresylic acids are fractionated to make o-, m-, and p-cresol, and xylenols, which are available for sale up to the market limit (10 percent of U.S. market).[9] If the cresylic acid fractionation system is built to handle the entire stream volume, the LP chooses to fractionate enough to meet the market limit on all stocks, even though this means dumping 6 BD excess xylenols to fuel. The arbitrary market limit is explored as a sensitivity later in this report.

Since the maximum profit case uses only the Great Plains naphtha and phenol streams, the tar oil stream is available to make jet fuels. The remaining cases use maximum available tar oil to make jet fuel. The profit in these cases is less than when no jet fuels are produced. Profitability for these cases depends on replacement fuel cost, which is reported later as a sensitivity.

Case 2 - Profitable JP-4.--The profitable JP-4 case is shown in Figure 16. The Great Plains tar oil is hydrotreated and hydrocracked as in Case 1. The Great Plains naphtha and phenol streams are processed to make chemicals and BTX, as in Case 7, but the amount of BTX extraction is adjusted to balance the octane requirements for blending with naphthas from the hydroprocessors to produce an unleaded gasoline product. The design basis for this case specifies aromatics recovery and cresylic acid fractionation units big enough to process the entire stream, since the incremental cost is small. As in the maximum profit case, a small volume of xylenols are diverted to fuel. Hydrogen for all three hydroprocessors is obtained from a PSA unit.

With the price structure assumed, it is better to sell some naphtha streams as reformer feed to existing refiners than to build a new reformer unit at the Great Plains site. Table 9 shows this sales option was selected in Cases 2, 4, and 6.

Case 4 - Profitable JP-8.--The profitable JP-8 case is shown in Figure 17. The Great Plains tar oil is hydrotreated and hydrocracked under almost identical conditions as in Case 3. The Great Plains naphtha and phenol

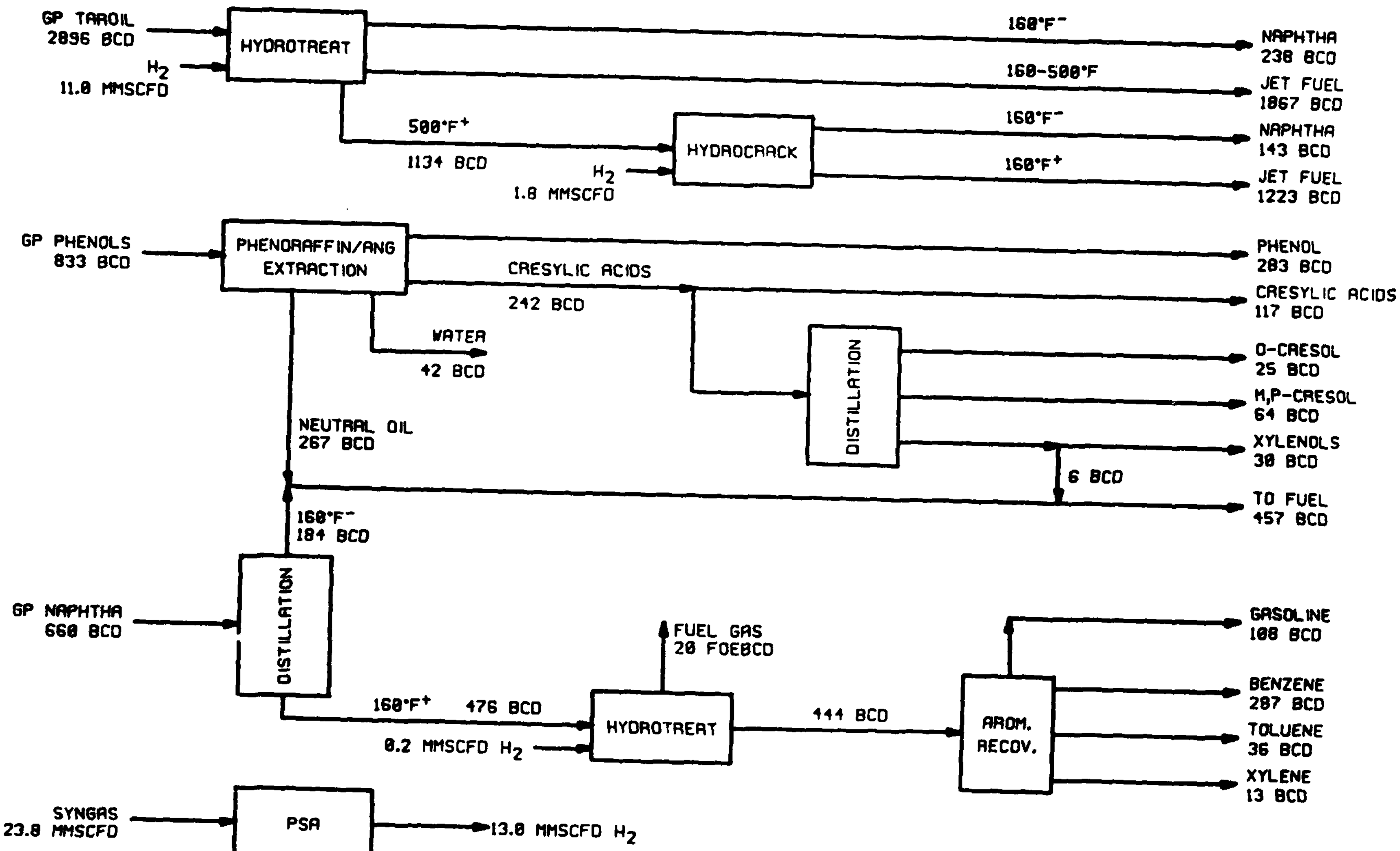

FIGURE 16
GREAT PLAINS LIQUID BY-PRODUCTS USE
CASE 2: PROFITABLE JP-4 CASE

TABLE 9

SUMMARY OF PROCESS DESIGN AND COST ESTIMATES

Case	1	2	3	4	6	7
Unit Production Rates (BPCD)*						
Jet Fuel	3698	3062	2864	2297	1744	-
Reformer Feed/Gasoline	-	437	-	1190	511	41
Phenol	-	285		285	285	285
Cresylic Acids	-	236	-	236	236	236
BTX	-	294	-	398	398	398
Capital Cost (THSD $)						
Jet Fuel/Naphtha	51,430	43008	44,799	42978	31111	-
Phenol/Cresylic Acids	-	20124	-	20124	20124	20124
BTX	-	15000	-	17664	17664	17664
TOTAL	51,430	78,132	44,799	80,766	68,899	37,788
Operating Costs ($/D)						
Utilities (Incl. SNG Equiv.)	79,230	89,980	51,130	90,947	74,416	21,121
Cat & Chem	2400	3416	2963	3331	3140	1373
STM Usage (#/H) HP	-	54,700	-	58,200	58,200	58,200
MP	5300(Exp)	8900	6945(Exp)	9,300	9,300	15,900
LP	-	6900(Exp)	-	7050(Exp)	7050(Exp)	6900(Exp)
Personnel	40	76	40	76	68	56

*Appendices give production rates based on barrels per stream day, not barrels per calendar day.

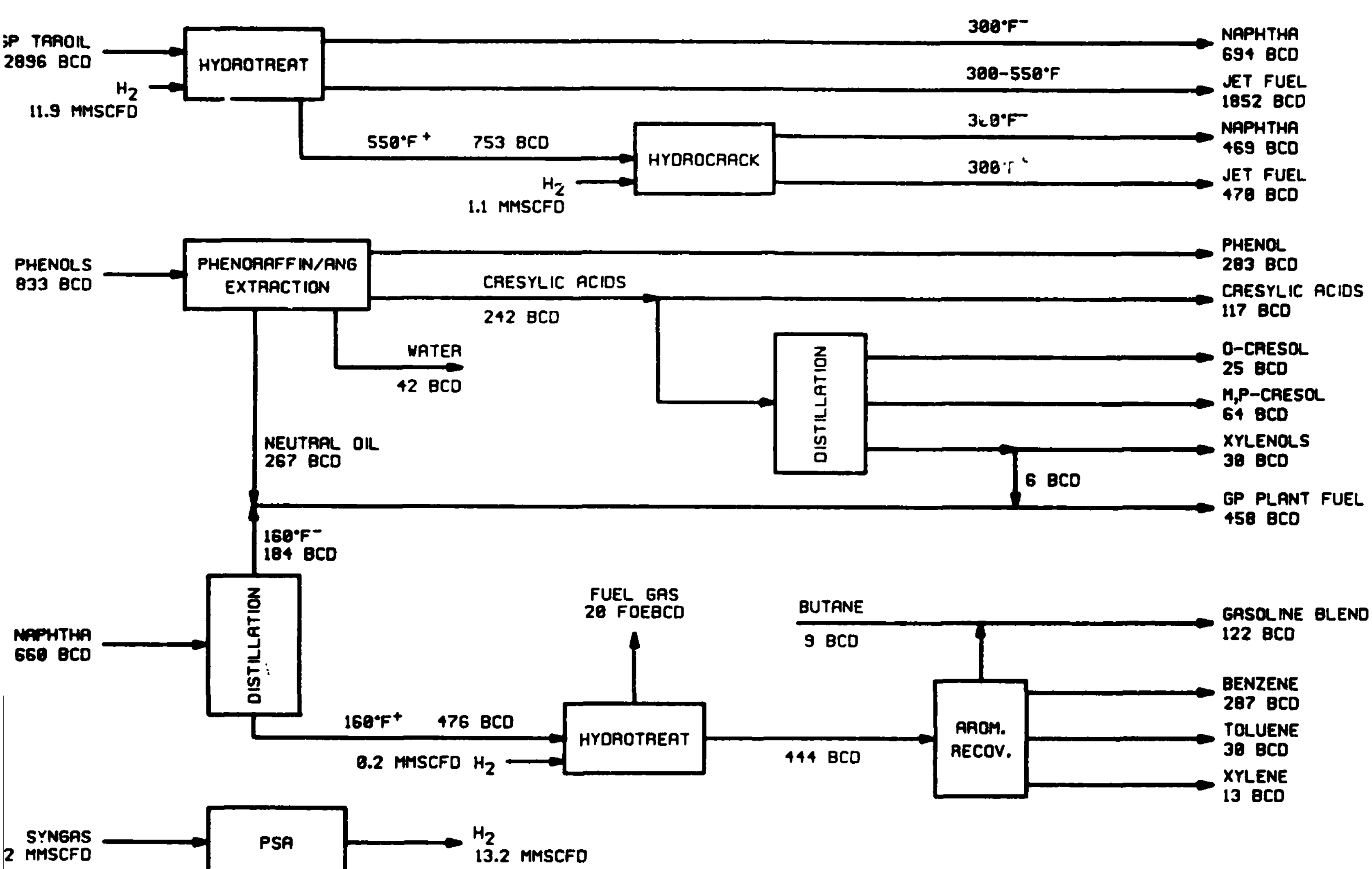

FIGURE 17
GREAT PLAINS LIQUID BY-PRODUCTS USE
CASE 4: PROFITABLE JP-8 CASE

streams are processed to make chemicals and BTX as in Case 7, but additional toluene is added to the gasoline blending stock in the aromatics recovery unit to balance the octane requirements for blending with naphthas from the hydroprocessors. The commercial design basis for this case would build the cresylic acid fractionation unit big enough to process the entire stream. Hydrogen for all three hydroprocessors is obtained from a PSA unit.

Case 6 - Profitable JP-8X.--The profitable JP-8X case is shown in Figure 18. To produce a volume of jet fuel comparable to Case 5, the LP hydrotreats the neutral oil from Phenoraffin extraction along with 85 percent of the 750°F- Great Plains tar oil. The remaining tar oil and the 750°F+ cut are used as fuel. The neutral oil stream, not available in Case 5, can be hydrotreated to produce JP-8X more cheaply than from tar oil. The Great Plains naphtha and phenol streams are processed to make chemicals and BTX, as in Case 7, but the amount of BTX extraction is adjusted to balance the octane requirements for blending with naphtha from the hydrotreater. The design basis for this case would build the aromatics recovery and cresylic acid fractionation units big enough to process the entire stream. Hydrogen for both hydroprocessors is obtained from a PSA unit.

A summary of the feeds and products for the LP Cases 1-7 are shown in Table 10, along with unit capacities for each case.

c. Process Design and Cost Estimate

Based on the LP conceptual process schemes, Lummus has developed conceptual process designs and cost estimates for the various product slates, with the exception of Case 5. Table 9 summarizes Lummus production rates, capital and operating costs, which have been developed for each case. Appendices B through H detail the results. Production rates in Tables 9 and 10 are different because Table 9 is based on a preliminary process design basis, whereas Table 10 gives rates which were re-optimized based on capital and operating costs from the preliminary process design and all the process and blending limits in the LP. Investments and utilities are also different in Tables 9 and 10 because of this re-optimization.

In agreement with Amoco and the DOE, the phenol stream process (Phenoraffin/ANG) design is based on data received from ANG Coal Gasification Co.[4] Consequently, the phenol extraction and cresylic acid distillation areas are not based on proven, licensable technology and remain the least certain processing blocks in the plant.

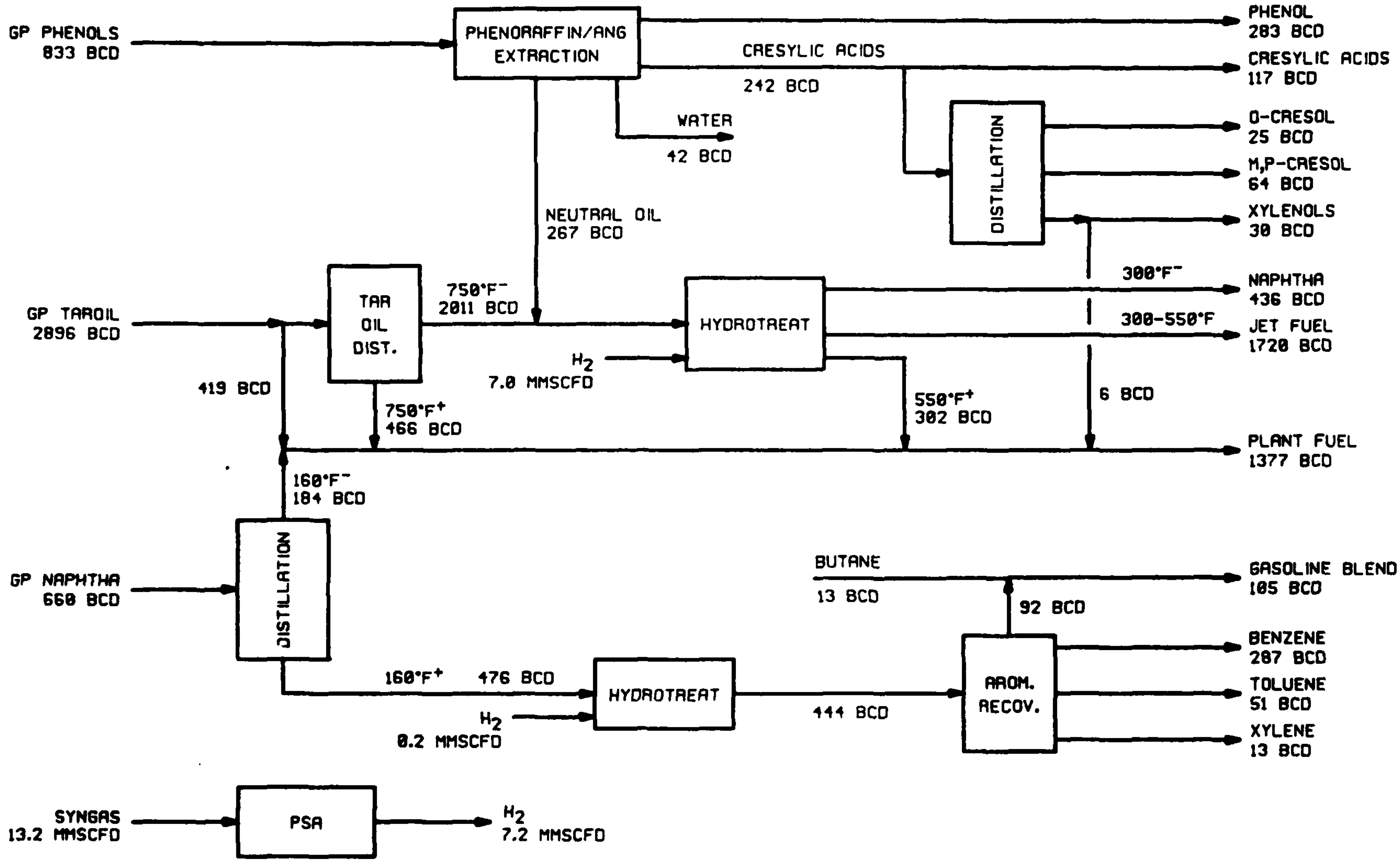

FIGURE 18
GREAT PLAINS LIQUID BY-PRODUCT USE
CASE 6: PROFITABLE JP-8X CASE

TABLE 10

GREAT PLAINS LIQUID BY-PRODUCTS CASE SUMMARY

(Total Upgrading Process)

Economics	Max Profit	JP-4 Max.	JP-4 Prof.	JP-8 Max	JP-8 Prof.	JP-8X Max	JP-8X Prof.
Profit, M$/CD	40.8	-3.2	32.7	-16.2	24.8	-14.7	29.0
Profit, MM$/yr	14.9	-1.2	11.9	-5.9	9.0	-5.3	10.6
Investment, MM$	37.0	53.1	87.1	52.7	89.3	33.7	68.5
Feedstocks, BCD							
G.P. Naphtha	660	0	660	0	660	0	660
G.P. Phenol	833	833	833	0	833	0	833
G.P. Tar Oil	0	2896	2896	2896	2896	2896	2896
Syngas, MMSCFD	0.4	30.4	23.8	23.7	24.2	14.3	13.2
Products, BCD							
Gasoline	62	466	336	239	297	0	165
Reformer Feed	0	0	154	983	988	379	376
Jet Fuel	0	4052	3090	2323	2322	1763	1720
BTX	389	0	336	0	330	0	351
Oxygenated Chem.	519	0	519	0	519	0	519
Liq. Fuel	457	0	457	0	454	947	1377
Unit Cap., BCD							
Aromatics Rec.	444	0	444	0	444	0	444
Phenoraffin/ANG	833	0	833	0	833	0	833
Tar Oil Dist.	0	0	0	0	0	841	2477
Hydrocrack.	0	1159	1134	753	754	0	0
Hydrotreat.	0	3729	3563	4797	4796	2379	1925
Naphtha Dist.	660	0	660	0	660	0	660
Naphtha Hydrotrt.	476	0	476	0	476	0	476
PSA, MMSCFD	0.4	30.4	23.8	23 7	24.2	14.3	13.2

As a result of the work accomplished during this task, two important items for the economic success of the by-product processing must be resolved.

The first and most apparent is the development of a process to provide a saleable cresylic acid product slate. Product specifications and limitations on production, if any, must be developed. Also, allowable levels of guaiacol and catechol in the cresylic acids must be addressed. One means of separating these close-boiling materials from the desired products may be through the use of crystallization. Although the compounds have close boiling points, their melting points differ substantially. It may be desirable to separate the components into narrow boiling point ranges and then purify each cut by fractional crystallization. Pilot plant work should be undertaken to develop an effective means of recovering these saleable products.

Secondly, Great Plains fuel requirements will increase significantly in some cases to replace the fuel presently being provided by the by-products. Since the replacement fuel required for the various profitable cases (1-4 MBD) is comparable to the residual fuel in LPG consumption of all North Dakota (3 MBD),[2] alternative fuel sources must be studied and incorporated into the design basis of these units.

3. Subtask 1.3 Economic Analysis

a. Assumptions and Basis

As is the case with all preliminary evaluations, a number of assumptions were made to reach the economic results which are presented in this section. Before presenting the results, a review of these assumptions is warranted.

In all seven cases,

1. It is assumed that the existing Great Plains plant can supply adequate high pressure steam and other utilities and power to the liquid by-product upgrading plant. Cost and availability information was supplied by ANG.

2. It is assumed that any H_2S generated by the upgrading facility can be processed to produce sulfur by the existing Great Plains sulfur plant. No capital or operating allowance has been made for a separate sulfur plant for by-product upgrading.

3. It is assumed that any ammonia and wastewater produced by the upgrading facility can be handled by the existing Great Plains ammonia recovery and wastewater treatment plants.

4. It is assumed that the existing boilers which currently burn the tar oil, naphtha, and phenolic streams can be modified to burn the selected alternative fuel with little or no capital investment. No capital has been included in the estimates for the various cases to modify the boilers. Furthermore, no allowance has been made for storage of the replacement fuel.

5. It is also assumed that the various processes which are considered by the LP can produce the required product at the commercial specifications for that product. As pointed out previously, this assumption is particularly important for the processes which handle the phenolic stream. Amoco has no commercial experience with these processes.

6. Numerous economic assumptions are made. It is assumed that:

   a. The upgrading plant is 100 percent equity financed.

   b. The upgrading facility is built in two years and started up in one year.

   c. A 10 percent real rate of return is required on invested capital.

   d. No land purchase is necessary.

   e. Upgrading plant life is 20 years and the plant has zero salvage value.

f. Federal and North Dakota state taxes combined are 40 percent of gross profit.

g. The plant onstream factor is assumed to be 0.91 for tar oil and naphtha processing and 0.89 for phenolic processing.

h. The upgrading plant is depreciated over a 10 year period using the Tax Recovery Act of 1986 schedule.

i. Various levels of investment contingency (10 to 30 percent) have been added according to process definition and maturity.

These assumptions are consistent with a capital charge factor of 16 percent/year.

Note that assumptions 1 through 4 will be dealt with during Task 4.

b. Economic Results

The conceptual process schemes to upgrade the by-products were developed by the Great Plains LP, based on Amoco estimates of capital and operating costs for the various process steps. As discussed above, Lummus developed detailed process designs and capital and operating costs for each conceptual case. The LP was revised with these Lummus cost figures and the economics of each case was estimated, assuming a replacement fuel cost of $2.15/MMBtu. In the discussion that follows, profit is above a 10 percent real rate of return on invested capital, as all costs include a capital charge on capital employed in the upgrading plant.

Table 11 shows the investment breakdown by unit. Table 12 summarizes the economics of the cases, broken down by cost component. All the maximum jet fuel cases show a loss (negative profit) because the total cost of upgrading

TABLE 11

GREAT PLAINS INVESTMENT AND UTILITIES SUMMARY

MM$	Max Profit	JP-4 Max.	JP-4 Prof.	JP-8 Max	JP-8 Prof.	JP-8X Max	JP-8X Prof.
Aromatics Rec.	12.3	0	12.3	0	12.3	0	12.3
Phenoraffin	19.4	0	19.4	0	19.4	0	19.4
Tar Oil Dist.	0	0	0	0	0	3.5	4.3
Hydrocrack.	0	13.2	13.0	11.2	11.2	0	0
Hydrotreat.	0	28.1	27.5	31.3	31.3	23.2	21.1
Naphtha Dist.	0.2	0	0.2	0	0.2	0	0.2
Naphtha Hydro.	4.5	0	4.5	0	4.5	0	4.5
PSA	0.5	10.5	8.8	8.8	9.0	6.2	5.9
Power Dist.	0.1	1.4	1.3	1.4	1.4	0.8	0.8
Total	37.0	53.1	87.1	52.7	89.3	33.7	68.5
Utilities (1)							
Cat & Chem, $/D	542	3533	3920	4470	5014	4389	4092
Fuel, FOEB/D	1068	4071	4330	3094	4336	2159	2935
Power, MW	0.2	6.2	5.8	6.0	6.3	3.2	2.9
Cool Wat., M gpm	2.9	2.2	4.9	2.3	5.2	1.0	3.8
Proc Wat., gpm	3	31	33	38	40	17	16

Note: Steam costs or credits are allocated in this table to fuel and cat & chem.

(1) These utilities do not agree with those in Table 9 because of reoptimization by the LP.

TABLE 12

GREAT PLAINS ECONOMICS SUMMARY

Cash Flow M$/CD	Max Profit	JP-4 Max.	JP-4 Prof.	JP-8 Max	JP-8 Prof.	JP-8X Max	JP-8X Prof.
Net Sales (1)	79.9	104.3	163.2	78.4	158.5	47.9	127.7
Fuel (2)	-14.6	-65.4	-66.7	-52.0	-66.9	-34.1	-48.3
Cat and Chem	-0.5	-3.5	-3.9	-4.5	-5.0	-4.4	-4.1
Utilities (3)	-0.9	-6.5	-6.5	-6.4	-7.3	-3.3	-3.7
MTIO (4)	-4.1	-5.8	-9.5	-5.8	-9.8	-3.7	-7.5
Fixed Costs (5)	-2.6	-2.8	-5.3	-2.7	-5.2	-2.3	-4.9
Capital Recov.(6)	-16.4	-23.5	-38.5	-23.3	-39.5	-14.9	-30.3
Total Profit	40.8	-3.2	32.7	-16.2	24.8	-14.7	29.0
Total MM$/yr	14.9	-1.2	11.9	-5.9	9.0	-5.3	10.6

Notes:

(1) Includes naphtha, gasoline, BTX, and chemicals, less the cost of purchased gasoline blending stocks (e.g., butane).
(2) Includes Great Plains naphtha, tar oil, phenol, and hydrogen removed from syngas, as well as purchased fuel, less credit for fuel returned to the Great Plains pool. Hydrogen is priced at a premium over fuel value. Replacement fuel is assumed to be LPG at $2.15/MMBtu.
(3) Includes power, process water, and cooling water. Steam costs and credits are allocated to fuel and catalysts and chemicals.
(4) Maintenance, taxes, insurance, and overhead charges.
(5) Primarily operating labor.
(6) Ten percent real rate of return, 5 percent inflation, 2-year construction, 1-year startup.

Jet Fuel, BCD Production Rate	0	4052	3090	2323	2322	1763	1720
Jet Fuel Unit Profit/Loss to Breakeven, $/B	-	-0.8	10.6	-7.0	10.7	-8.4	16.9

the feeds to jet fuels and of replacement fuel exceed the revenue generated by jet fuel sales. Table 6 shows that oxygenated chemicals and BTX are valued high enough above feed prices that cases which make these products from the Great Plains phenol and naphtha streams generate a profit. The maximum profit case could potentially add about 15 MM $/yr to the Great Plains annual profits. Profitability of the cases which make jet fuels, besides chemicals and BTX, are between 9 and 12 MM $/yr.

Another way of looking at the economics is to calculate how much the jet fuel price would have to be subsidized to break even for the maximum jet fuels cases, and how much the jet fuel price could drop to break even for the profitable jet fuels cases. These values, shown in Table 12, suggest that a subsidy less than a $9/B is needed to break even for the maximum jet fuels cases; in fact, the JP-4 case subsidy is less than $1/B. If BTX and chemicals by-products are produced, jet fuel sales prices could be lowered by $10-17/B and still break even.

c. Sensitivities

The economics discussed above are affected by several key assumptions. The effects of altering these assumptions is explored in this section. The objective of this section is to provide a scoping estimate of the impact of these assumptions on the economics. Lummus has not confirmed Amoco's estimated capital and operating costs for some of the process schemes discussed in this section.

The economics presented above assume that the Great Plains stocks could be replaced with other fuel stocks at $2.15/MMBtu. This value is roughly the current cost of LPG in the North Dakota area. Table 13 shows the values of LP streams if other fuel costs are assumed from $1.5C/MMBtu to $5.00/MMBtu, by adjusting for the Btu content of the various streams. The LP has been optimized with these values, and the effect of fuel price on profitability for the maximum jet fuels cases is shown in Figure 19. Higher fuel costs have a strong negative effect on profit. Fuel costs at least below $2.00/MMBtu are

TABLE 13

EFFECT OF FUEL PRICES ON STREAM VALUES

Stream	Unit	Fuel Price, $/ MM Btu 1.50	2.15	3.80	5.00
Natural Gas	FOE B	9.45	13.57	23.94	31.50
LPG	Bbl	5.26	7.57	13.32	17.53
n-Butane	Bbl	11.76*	11.76*	15.16	19.95
Grt Pl Naphtha	Bbl	8.85	12.69	22.42	29.50
Grt Pl Phenol	Bbl	7.28	10.43	18.43	24.25
Grt Pl Tar Oil	Bbl	9.12	13.07	23.10	30.40
Syngas for H2	MSCF	0.67*	0.67*	1.08	1.42
Utilities Fuel	MMBtu	1.50	2.15	3.80	5.00

* Current price is above fuel value. Current price is used.

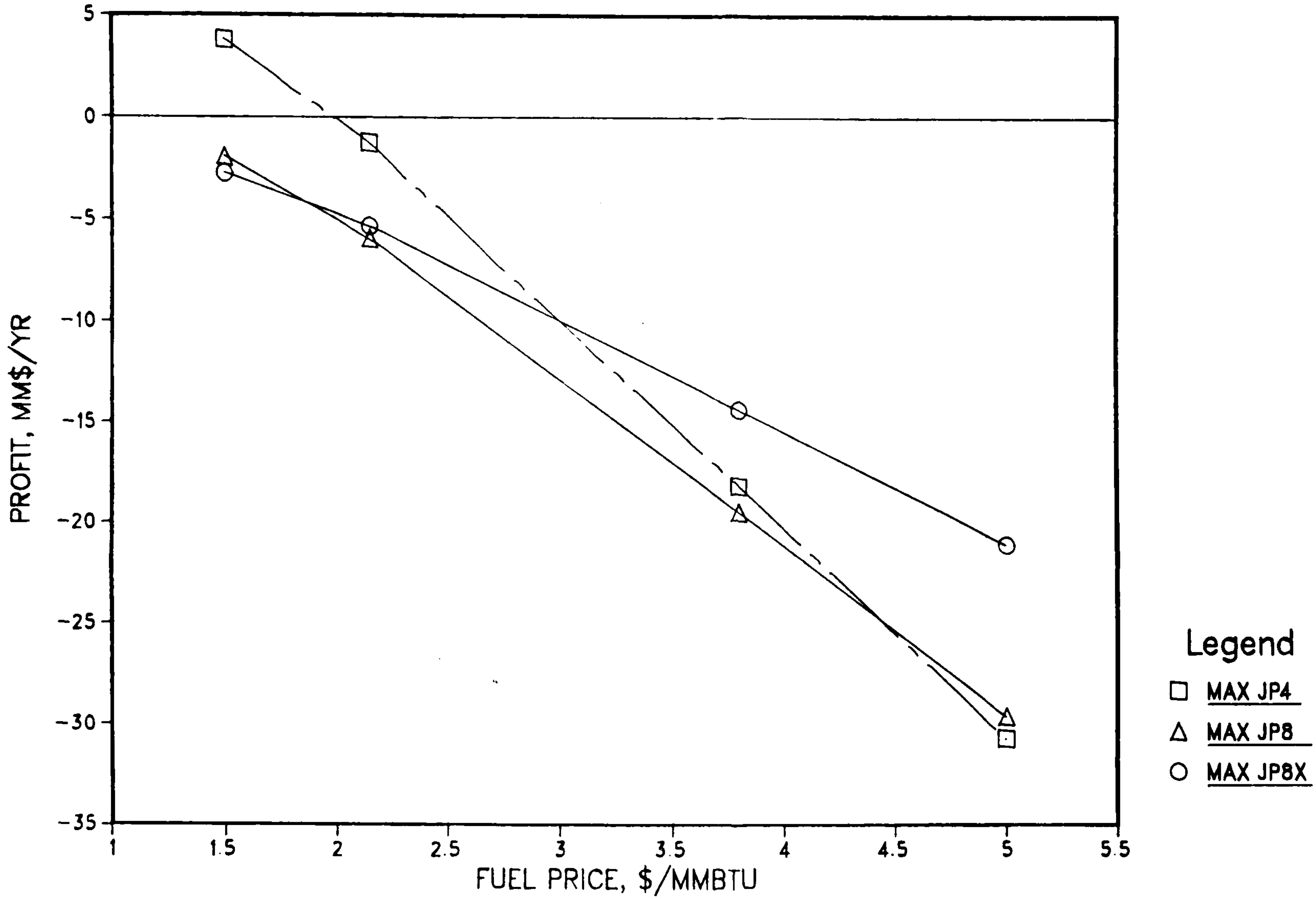

FIGURE 19
MAXIMUM JET FUEL CASES
EFFECT OF FUEL PRICE

needed to generate a profit in the maximum JP-4 case. Fuel costs below $1.50 MMBtu would be required to generate a profit in the JP-8 and JP-8X cases.

Figure 20 shows the effect of varying product slates along with fuel price. The maximum profit case retains profitability for all fuel values studied, although profit falls by about $2 MM/yr for each $1/MMBtu increase in fuel costs. The maximum JP-4 case is copied from Figure 19. If the phenolic stream in the maximum JP-4 case is processed to make oxygenated chemicals while processing the tar oil to make JP-4, the profitability is increased about $10 MM/yr. This processing is profitable if fuel costs are below about $3.10/MMBtu. If, in addition, the Great Plains naphtha stream is processed for BTX, profits increase further at low fuel costs. Break even fuel cost is about $3.20/MMBtu for this processing.

Figure 21 shows the effect of fuel cost on profitability for the profitable jet fuels cases. The JP-8 case is profitable for all fuel costs below $3.00/MMBtu; the JP-4 case is profitable below $3.20/MMBtu; and the JP-8X case is profitable below $3.50/MMBtu. Profits are below the case which produces no jet fuels, and these cases are also more sensitive to fuel costs.

d. Hydrogen Cost

Table 14 shows that the hydrogen cost component in the maximum profit case is negligible, so that case would not be affected by hydrogen cost variation. The cases which produce jet fuels have hydrogen costs between 3 and 7 MM$/yr. Doubling hydrogen costs would therefore decrease profits by 3 to 7 MM$/yr.

e. Effect of Oxygenated Chemical Market Size

Figure 22 shows the effect of imposing various limits on the sales of cresols, xylenols, and cresylic acids as a percent of the U.S. market.[9] The starting basis for this discussion is Case 7 at 10 percent of the U.S. market for these oxygenated chemicals. As Figure 15 shows, in Case 7 the Great

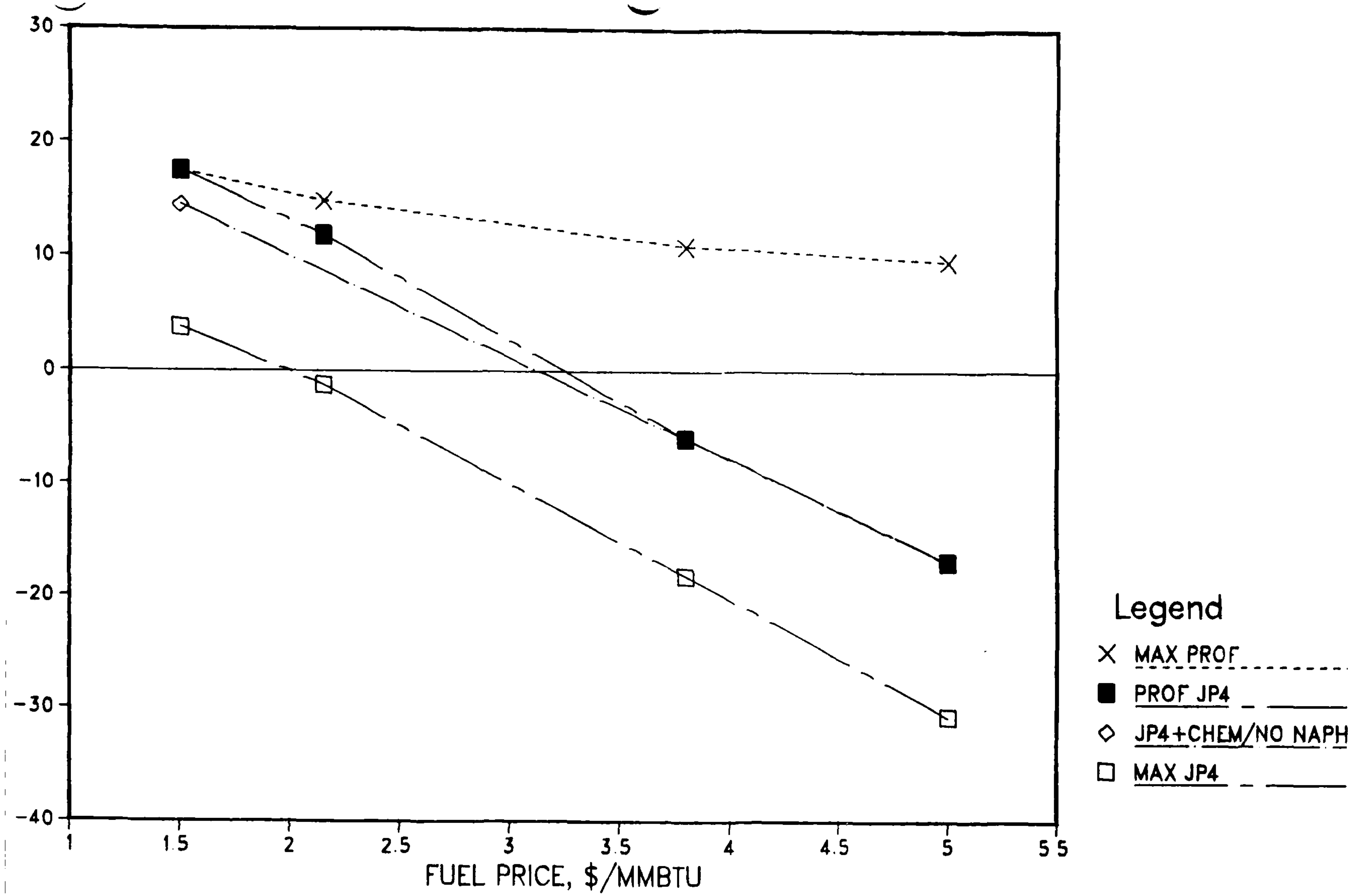

FIGURE 20
CASES WITH JP-4
EFFECT OF FUEL PRICE

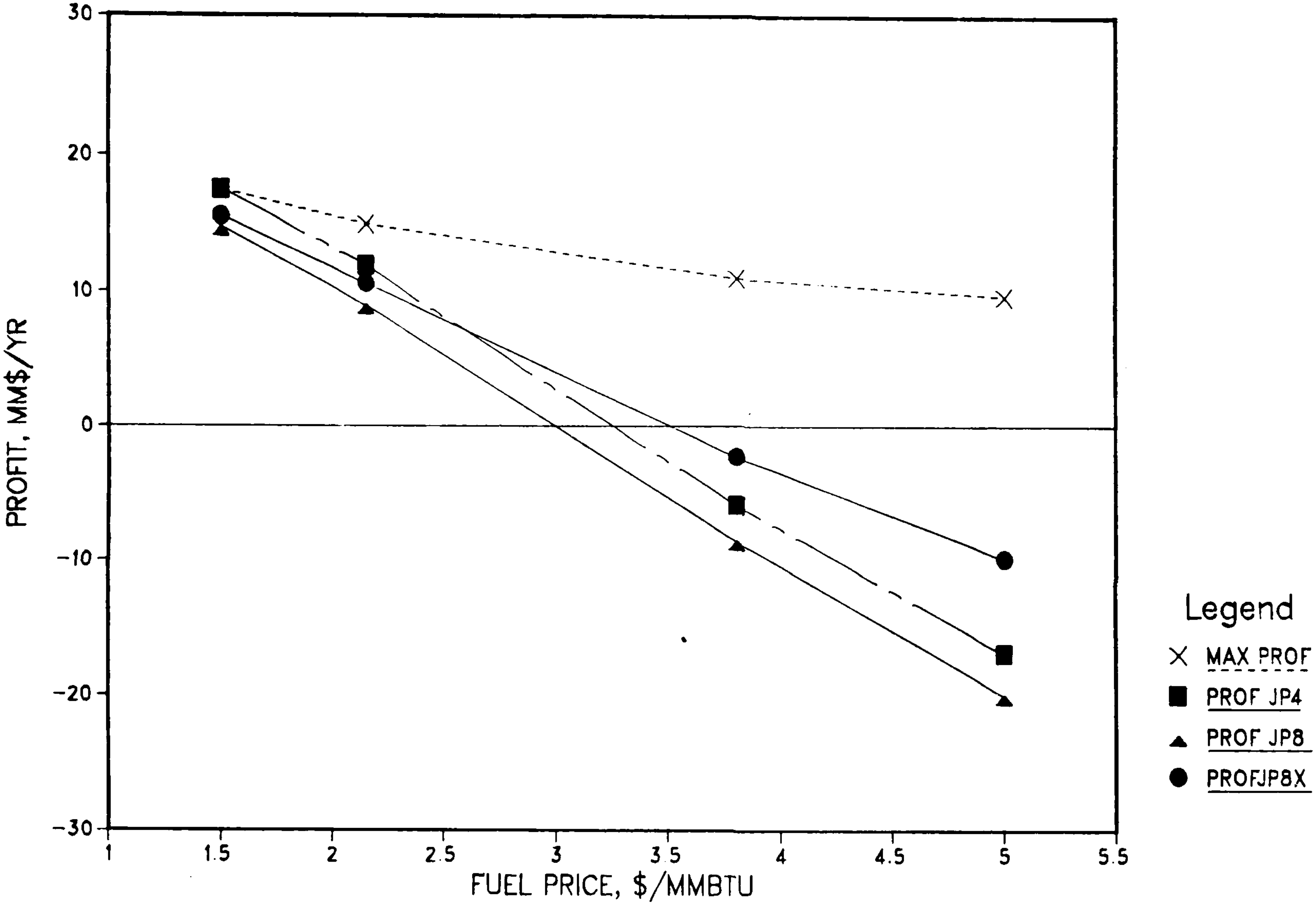

FIGURE 21
PROFITABLE CASES
EFFECT OF FUEL PRICE

TABLE 14

EFFECT OF HYDROGEN COST ON PROFITABILITY

Case	Syngas, MMSCFD	Cost* M$/CD	Cost* MM$/yr
Maximum Profit	0.4	0.2	0.1
Maximum JP-4	30.4	20.4	7.4
Maximum JP-8	23.7	15.9	5.8
Maximum JP-8X	14.3	9.6	3.5
Profitable JP-4	23.8	15.9	5.8
Profitable JP-8	24.2	16.2	5.9
Profitable JP-8X	15.2	8.8	3.2

* at $1.23/MSCF of hydrogen removed.

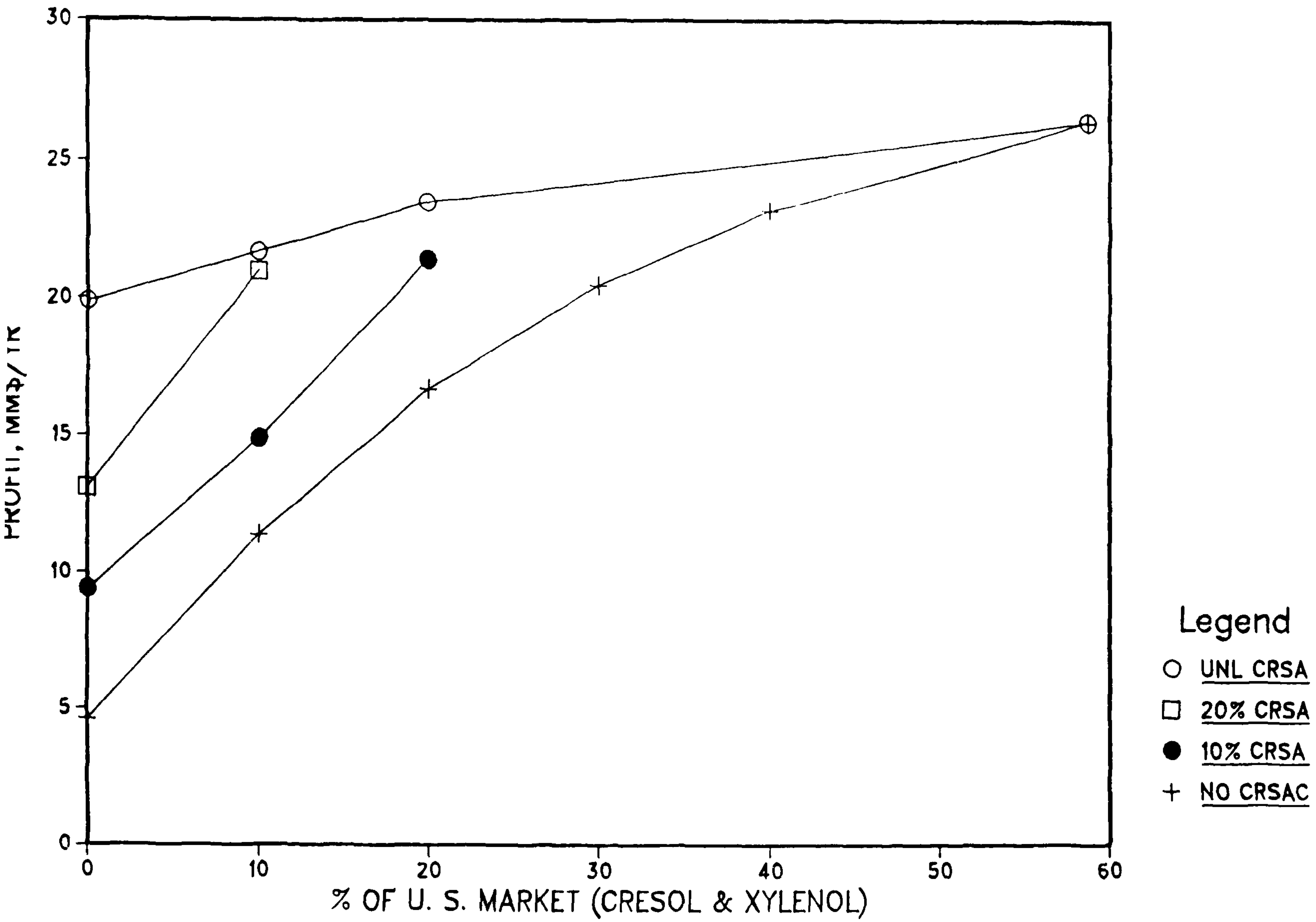

FIGURE 22
MAXIMUM PROFIT CASE
EFFECT OF CHEMICALS SALE

Plains phenols stream is extracted to make phenol and cresylic acids. Then some cresylic acids are fractionated to make maximum amounts of cresols and xylenol. Remaining cresylic acids are sold.

If cresylic acids sales are limited to lower levels, a Dynaphen unit is built to convert any non-fractionated cresylic acids into benzene and phenol. Fractionation for sales as cresol and xylenol is always favored over cresylic acid sales or Dynaphen processing. Figure 23 shows the process units when Dynaphen is maximally used (at 0 percent of cresol, xylenol, and cresylic acid sales). In this process scheme, all cresylic acids from extraction are processed in Dynaphen. It may be possible to process the Great Plains phenol stream directly with Dynaphen without prior extraction by distilling phenol from the cresylic acid. This would improve the economics of this case by about $4 MM/yr in capital and operating costs. All points in the lower left of Figure 22 involve some Dynaphen processing, as illustrated by Figure 24.

As sales volumes are increased from 10 percent of the U.S. market, cresol and xylenol sales are satisfied by fractionating additional cresylic acids. When cresol and xylenol and cresylic acid volumes combined reach or exceed 30 percent of the U. S. market, the tar oil stream is fractionated so that additional chemicals can be extracted from 450°F- tar oil. Ultimately, all 450°F- tar oil is extracted, and all cresylic acids from both streams are fractionated into cresols and xylenols. Figure 22 shows that this is reached at about 60 percent of the U.S. xylenols market, although cresols volume are feedstock limited to 25 to 35 percent of the U.S. market. The change of slope in the lines observed in Figure 22 at market size greater than 20 percent of U.S. market is due to the fact that the 450°F- tar oil is leaner in oxygenated chemicals than the phenolic stream. Also shown in Figure 24 are the estimated regions of Dynaphen and 450°F- tar oil processing.

These economics assume the sales prices listed in Table 6. It is doubtful if these prices would exist if Great Plains significantly expanded the U. S. supply without further market development.

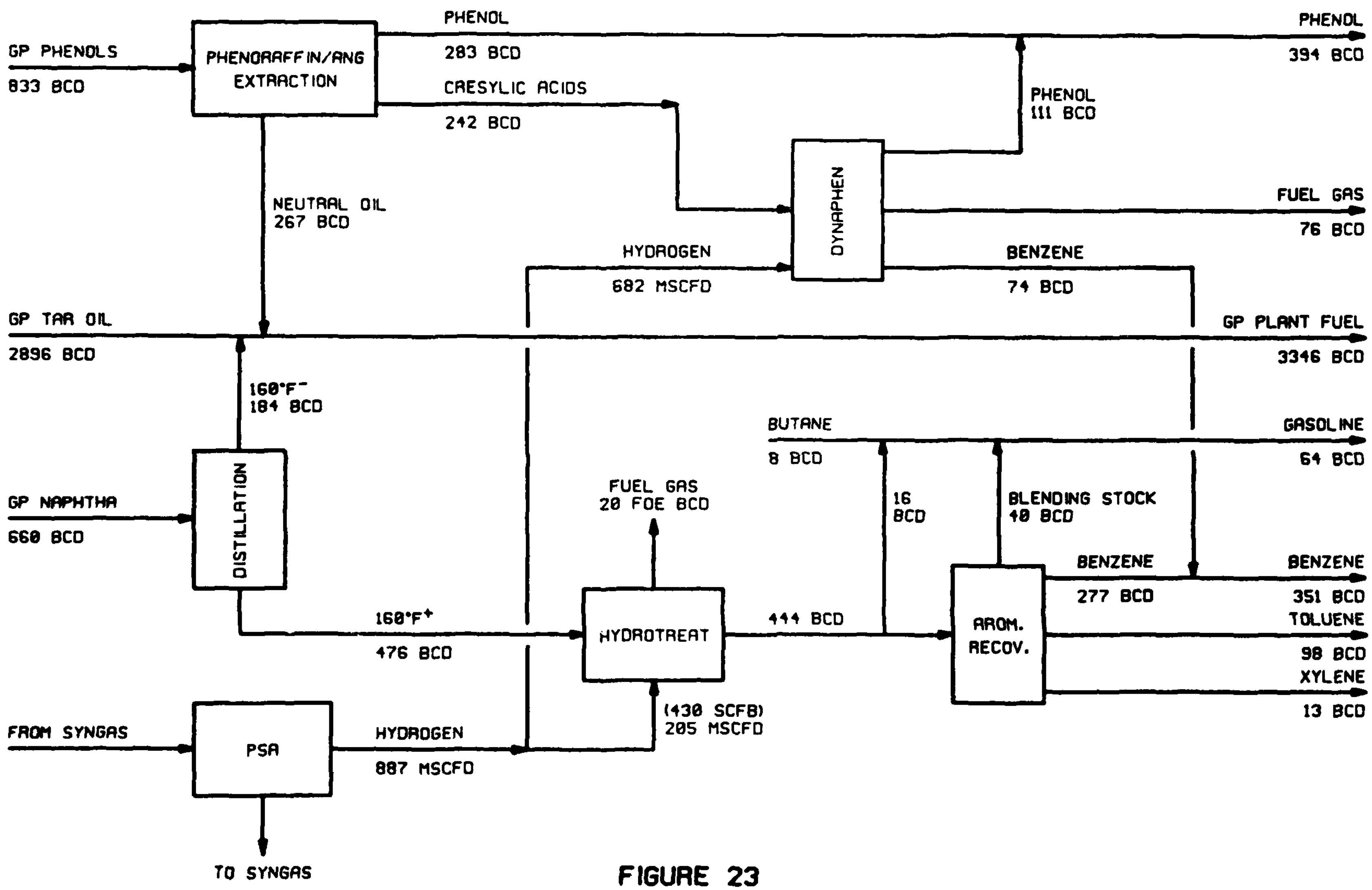

FIGURE 23
GREAT PLAINS LIQUID BY-PRODUCTS USE
WITH NO CRESOLS OR CRESYLIC ACIDS

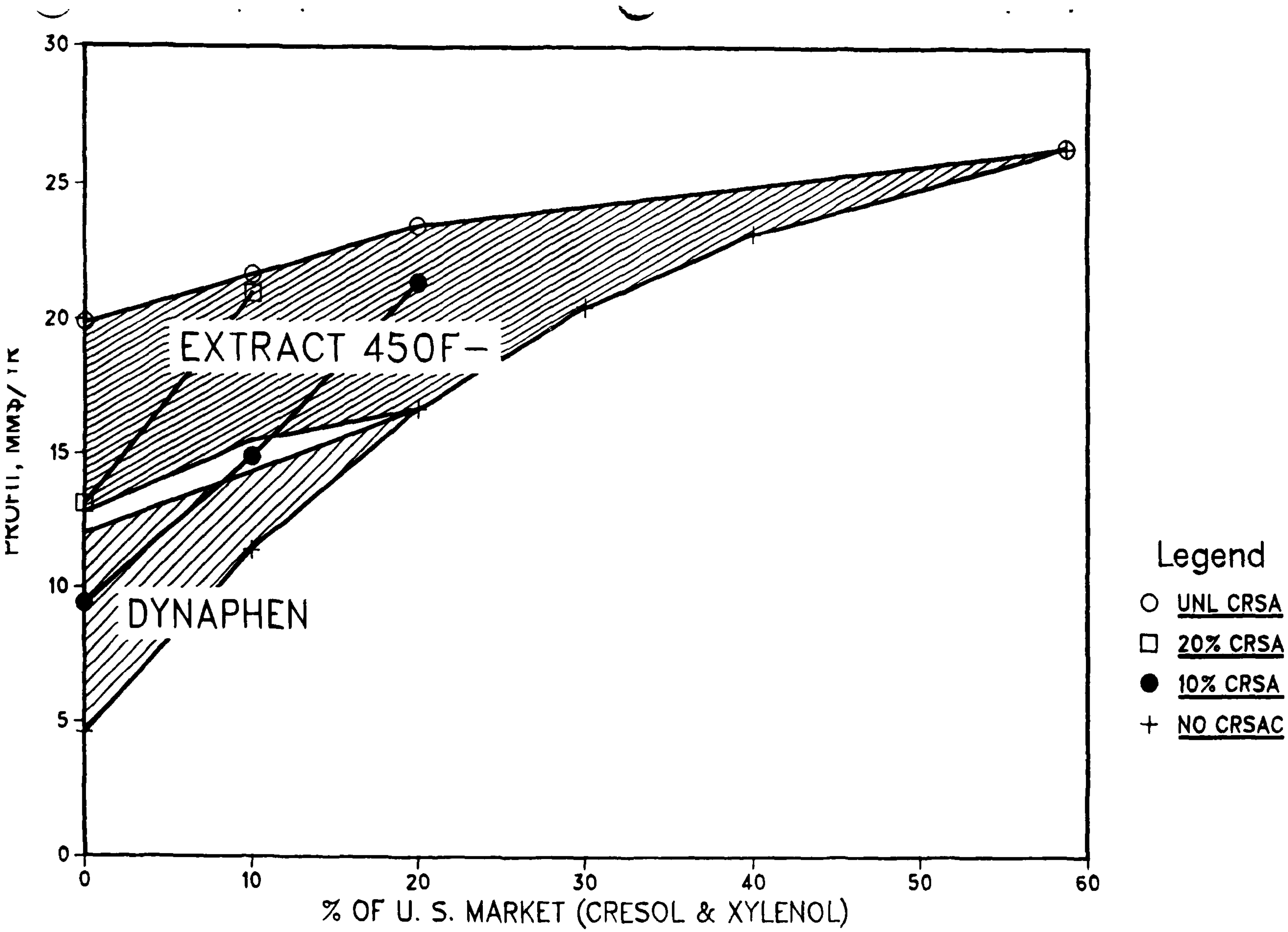

FIGURE 24
MAXIMUM PROFIT CASE
CHEM PROCESS ALTERNATIVES

SECTION V

SUMMARY AND CONCLUSIONS

A preliminary evaluation of the upgrading alternatives for the by-product streams (naphtha, phenols, and tar oil) from the Great Plains Gasification plant has been completed. The evaluation is based on (1) Amoco and other laboratory analytical analyses of the three streams, (2) the Western Research Institute scoping studies on hydrotreating the tar oil, (3) a market analysis of the various products by J. E. Sinor Consultants, Inc., (4) Amoco's proprietary process models for various refining technologies, (5) Lummus's cost estimates for the various required processes and (6) literature and ANG information on processes to upgrade and separate the phenolics stream. As a result of the analyses, seven possible upgrading schemes for the by-product streams were developed. The product slates in these schemes ranged from the maximum production of the various grades of jet fuel (JP-4, JP-8, JP-8X) to a slate which produced various chemicals (cresylic acid, phenol, cresols, xylenol, benzene, toluene, and xylene). As a result of this preliminary evaluation, the following conclusions have been reached:

1. The aging of the various by-product streams under storage conditions is minimal. Thus the quality of the by-product streams will not deteriorate in intermediate storage.

2. The quality of the by-product streams produced from the gasifier does not vary greatly with time. Thus the by-product quality bases used in our conceptual designs are representative.

3. The various grades of jet fuel can be produced from the tar oil at Great Plains, but not economically.

4. The phenolic and naphtha streams have the potential to significantly increase revenues at Great Plains.

5. The phenolic and naphtha streams could provide sufficient revenue to subsidize the production of jet fuel.

6. The economics of jet fuel production are sensitive to replacement fuel cost.

7. The amount of cresols and cresylic acid which can be marketed is a concern.

8. Processing the phenolic stream to produce cresols and cresylic acid should be demonstrated. The potential problems with contaminants such as guaiacol, catechol, and other contaminants should be assessed.

SECTION VI

RECOMMENDATIONS

An important goal of Task 1 is to recommend experiments which would be carried out in Task 2 to confirm the major (significant) assumptions of Task 1. Recommended experiments are:

For the tar oil by-product:

1. Additional hydrotreating data are needed to confirm yields, upgraded tar oil qualities, process conditions, and hydrogen consumption. These variables have the greatest impact on the economics.

2. Heat release as a function of product quality should be estimated to confirm the choice of an expanded bed over a fixed bed hydrotreater.

3. The impact on hydrotreating requirements of blending the phenolic stream with the tar oil stream to maximize JP-4 production should be determined.

4. The effect of catalyst type on JP-8X properties should be determined. (Can a catalyst be found which will allow saturation of sufficient aromatic rings, without opening the rings, so that the JP-8X specifications on aromatics and paraffin content can be met?).

5. The 500°F+ and the 550°F+ fractions of the hydrotreated tar oil should be hydrocracked to determine yield, product quality, process conditions, hydrogen consumption and heat release.

For the naphtha stream:

--The naphtha stream should be hydrotreated to determine yields and product quality (mainly heteroatom content of hydrotreated naphtha).

For the phenolic stream:

--The Phenoraffin or equivalent process should be demonstrated with the Great Plains phenolics stream. The yields and qualities (purity) of the various chemicals produced should be determined. Since this is potentially the most profitable stream to process, it is important that DOE or ANG do this. Evaluation of these processes is currently beyond Amoco's experimental capabilities.

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APPENDIX A

SATURATION KINETICS

APPENDIX A
SATURATION KINETICS

The tar oil, which at 3,180 b/sd is the largest volume liquid stream produced at Great Plains, has the greatest potential as a feedstock for jet fuel production. The very low H/C ratio and API gravity of this stream reveal that it is highly aromatic. Indeed, GC/MS analyses of the tar oil by Wilson and co-workers at the University of North Dakota Energy Research Center (UNDERC) have shown that it contains 90-95% aromatics, with the rest being paraffins.(3) Hydrogenation data from WRI(2,4) indicate that significant amounts of specification jet fuels can be made from the tar oil, but at the cost of very severe processing conditions (≥ 2,000 psig total pressure) and high hydrogen consumption (3,000-4,000 SCF/bbl). Much of this hydrogen is consumed in order to meet the jet fuel specification for aromatics content (<25%).

Aromatics Saturation Kinetics

As indicated by the above discussion, a central unit for processing the tar oil and crude phenols to make jet fuels would be a hydrotreater. The WRI data indicate that the limiting factor in this hydrotreater would be saturation of the tar oil aromatics. Thus, the kinetics of tar oil aromatics saturation play an important role in the design of this hydrotreater.

The WRI hydrogenation data with the whole tar oil(4) and Chevron hydrogenation data with SRC-II liquids(5,6) were used to develop a rate expression for tar oil aromatics saturation. The SRC-II data were chosen because these liquids resemble the Great Plains tar oil in terms of heteroatom content, aromatics content, and, to a lesser extent, boiling range. The rate expression resulting from this analysis is:

$$-r_A = k_o \exp(-E_A/RT)(P_{H_2})^{1.45}(1-X_A)^{1/2}$$

where:

X_A = fractional conversion of aromatics
R = gas constant
r_A = rate of aromatics saturation, hr^{-1}
k_0 = rate constant = f(feed, catalyst)
E_A = activation energy = 23,700 Btu/lb-mol
T = average reactor temperature, °R
P_{H_2} = average reactor hydrogen pressure, psia

This expression indicates that the rate of aromatics saturation is half-order with respect to the concentration of aromatics.

APPENDIX B

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 1
MAXIMUM JP-4 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
JAN 30, 1988

TABLE OF CONTENTS

1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to maximize the production of type JP-4 aviation turbine fuel from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- Two by product streams are combined and charged to the hydrotreater (Area 100).

    - Tar Oil 47620 #/hr (3182 BPSD)
    - Phenol 14213 #/hr (920 BPSD)

- The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 500^{o}F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (3850 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.

- The hydrotreater produces 6 streams:

    - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

    - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

    - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to the main boiler in the SNG plant.

    - JP4 turbine fuel which is combined with JP4 produced in the hydrocracker (area 200) and sent to storage.

    - 500^{o}F+ unconverted bottoms product which is sent to the hydrocracker (area 200).

    - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

- Approximately 750 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.

. The 500^{o}F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-4 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.

. The hydrocracker produces 4 streams in addition to JP-4.

- High pressure purge as (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H_2 and CH_4.

- Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

- Stabilized naphtha which is sent to the main boiler in the SNG plant.

- A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.

. Hydrogen make-up for both the Hydrotreater and the Hydrocracker is supplied from a PSA Hydrogen Unit. High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available at low pressure (5 psig) which has a fuel value of about 565 BTU/ft^3. This H_2, CO & CH_4 rich gas is recompressed into the methanation unit of the SNG plant.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. Detailed material balances for each unit can be found in appendixes A&B. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas and naphtha produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

4109 BPSD of Tar Oil and Phenol Feed
3372 BPSD of #6 Fuel Oil
13.34 MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

4278 BPSD of JP-4 turbine fuel
8.53 MMSCFD Equivalent SNG product credit due to HDT, HDC & PSA purge gas reinjection into SNG Plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	3372 BPSD
SNG Equivalent of Syn Gas & Purge Gas	4.81 MM SCFD
Power	7100 kW
Cooling Water	2400 GPM (30^{o}F rise)
Process Water	18.5 GPM

In addition the process exports 5280 #/hr of 100 psig saturated steam which was credited against boiler requirements.

Figure 1:Case 1-Maximum JP-4 Production

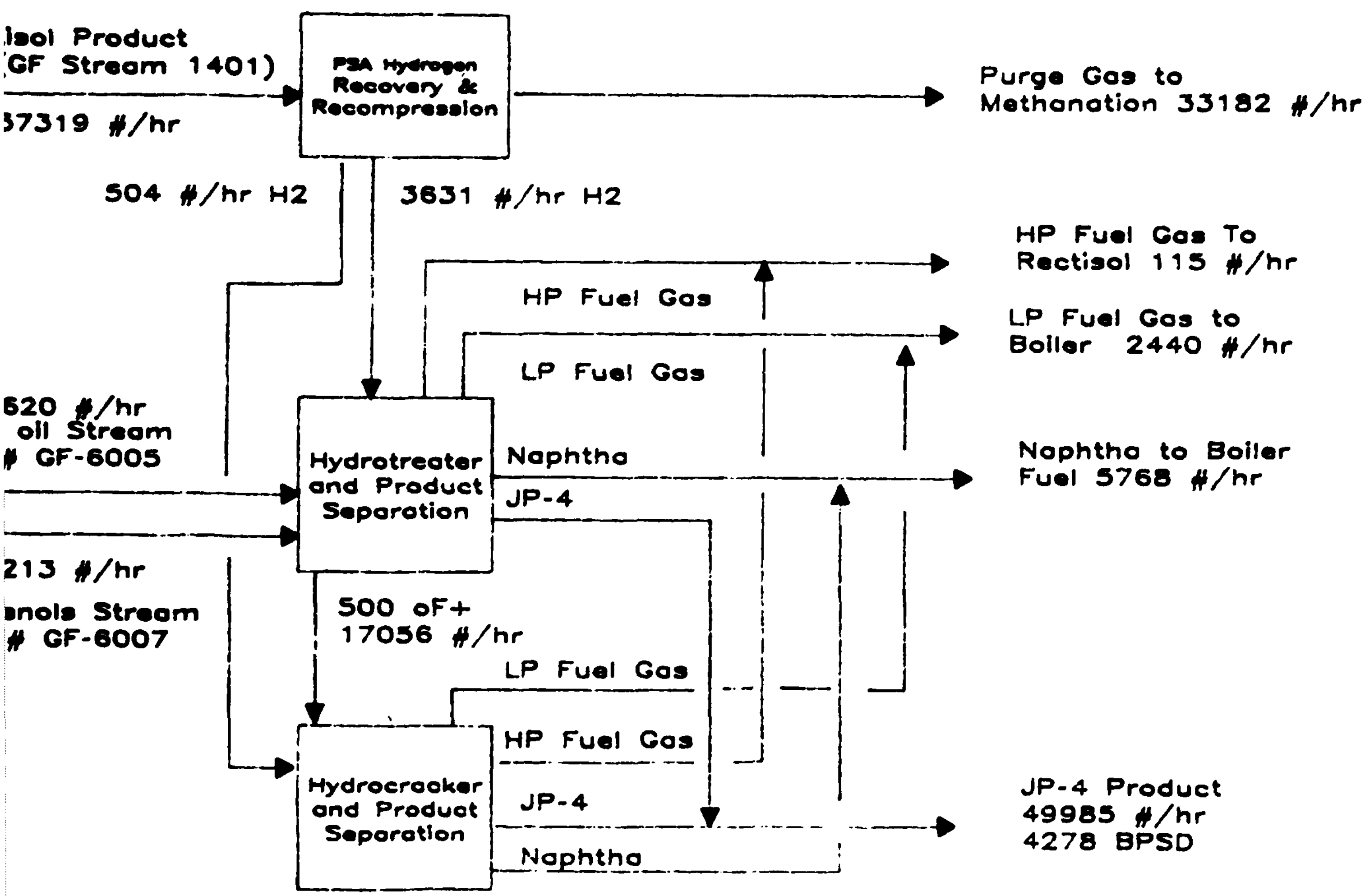


Table 1.1 Great Plains Case 1: Maximum JP4 Production

Tar Oil Feed=====>	47620	#/hr	3182	BPSD
Phenol Feed======>	14213	#/hr	920	BPSD
Total Oil Feed===>	61833	#/hr	4102	BPSD
JP-4 Product=====>	49985	#/hr	4276	BPSD
SNG Product Loss=>	8268	#/hr	4.8	MMSCFD
Fuel Oil Makeup==>	46646	#/hr	3372	BPSD

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feeds					
H2	5.87		3631	1801.1	
Oil	100.00	1.0342	61833		4102
Total	105.87		65464		
Products					
Purge Gas	0.13		79	24.5	
Fuel Gas	2.13		1319	65.0	
Naphtha	5.50	0.7238	3403		323
JP-4	58.14	0.8203	35948		3005
500 oF+	27.58	0.9465	17056		1236
H2O in SW	11.06		6839	379.9	
H2S in SW	0.35		218	6.4	
NH3 in SW	0.97		602	35.4	
Total	105.87		65464		4564

Fixed Bed Hydrocracker

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feeds					
H2	2.95		504	250.0	
500oF+	100.00	0.9465	17056		1236
Total	102.95		17560		
Products					
Purge Gas	0.21		36	10.3	
Fuel Gas	3.95		673	24.0	
Naphtha	16.49	0.6787	2813		284
JP-4	82.30	0.7563	14037		1273
H2S in SW	0.003		0.5	0.02	
NH3 in SW	0.003		0.5	0.03	
Total	102.96		17560		1558

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
HDT Nap	54.75	0.7238	3403		323
HCR Nap	45.25	0.6787	2813		284

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	2385.0	702.7	55.8	612.0	11.9	7.2	3774.6
Prod. H2	2051.1	0.2	0.0	0.0	0.0	0.0	2051.3
Purge Gas	333.9	702.5	55.8	612.0	11.9	7.2	1723.3
#/hr							
Feed Gas	4808	19681	2458	9817	357	229	37352
Prod. H2	4135	5	0	0	0	0	4140
Purge Gas	673	19676	2458	9819	357	229	33212

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	1319	65.0	25.1
HCR FG Produced	673	24.0	12.8
Stab FG Produced	448	9.8	8.2
Total Fuel Gas	2440	98.9	46.0

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	673	333.9	324	41.0
CO	19676	702.5	321	85.5
CO2	2458	55.8	0	0.0
C1	9819	612.0	1010	234.3
C2	357	11.9	1769	8.0
N2+Ar	229	7.2	0	0.0
Total	33212	1723.3	565	368.7

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Crude Phenol	-14213	13070	-185.8			-920
Tar Oil	-47620	17000	-809.5			-3182
Fuel Gas	2440	18853	46.0	0.9	1228	
Naphtha	5768	20040	115.6			574
Export Steam	5280	1000	5.3			
Fuel Oil to Boiler	46024	18000	828.4			3327
Total	-2321		0.0	0.9		-201
Fuel Oil to Process Heaters	622	18000	11.2			45

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	13.34	90591
SNG Credit for PSA Purge	8.40	41360
SNG Credit for Hdtrs purge gas	0.13	836
Total SNG Production Loss	4.81	

Water Balance Hydrotreater

Component	H2O (#/hr)	H2S	NH3	NH4HS	NH4OH	Total
Reaction Gases	6839	218	602			7659
Reaction Solution	6317			327	1015	7659
Stripping Steam	1856					1856
Softened Water	6826					6826
HDT Sour Water	14999			327	1015	16340

Water Balance Hydrocracker

Component	H2O (#/hr)	H2S	NH3	NH4HS	NH4OH	Total
Reaction Gases		0.5	0.5			1.0
Reaction Solution				0.8	0.5	1.3
Stripping Steam	556.9					556.9
HCR Sour Water	556.9			0.8	0.5	558.2

Total Sour Water

	H2O (#/hr)	H2S	NH3	NH4HS	NH4OH	Total
Total Sour Water	15556			328	1015	16898

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

Operating conditions for the hydrotreater were provided to Lummus by Amoco and these conditions are presented in Table 2.1. The basic processing step selected was the expanded bed hydrotreater (LC Fining) system. Due to the extremely high exothermic heat of reaction it was necessary to use 3 reactors in with interstage cooling. Referring to drawing D5571-10101 and the material balance printouts (Appendix A) the flow is as follows:

- Feed Tar Oil and Phenol are charged into the hydrotreater from day tanks FA-101 and FA-102 through charge pumps GA-101 and GA-102 and preheater exchanger EA-101.
- The preheated charge oil is combined with feed hydrogen gas (at 576^oF) from heater BA-101. Preheat of the oil is limited to 505^oF to prevent cracking. The preheated mixture is then charged to the first reactor DC-101A.
- The expanded bed reactor DC-101A approaches isothermal conditions in which the heat of reaction is used to heat the feed up to 700^oF.
- The effluent from DC101A is cooled in exchanger EA-101 and combined with recirculating hydrogen from recycle hydrogen gas compressor GB-102. The combined mixture (which has been cooled to 486^oF) is charged into the second reactor where the heat of reaction increases the temperatures to 700^oF.
- The effluent from DC101B is cooled in exchanger EA-104 and combined with recirculating hydrogen gas from recirculating compressor GB-102. This mixture is charged into the third reactor DC-101C where its temperature rises from 531^oF to 700^oF.
- The effluent from DC101C goes to the high temperature/high pressure separator FA-103. Hot liquid from FA-103 flows to the hydrotreator fractionation DA-101. The vapors from FA-103 flows through exchangers EA-102 and EA-105 and then through air cooler EC-101. Process water is injected prior to EC-101 to convert the H_2S and NH_3 in the gas to an aqueous NH_4OH/NH_4HS solution.
- Exchangers EA-104 and EA-105 are part of a circulating hot oil belt which allows for the generation of steam from waste heat in the high pressure loop without having the problem of a hydrocarbon leak from the high pressure system into the steam system.

2.1 Hydrotreater - Cont'd

- The cooled gas then passes into the High Pressure/Low Temperature Separator FA-104 where hydrogen rich gas is taken as an overhead product. A purge stream of this high pressure gas is taken (to purge H2 and light gases from the loop) and sent to the Rectisol Unit 1400 in the SNG plant to recover the hydrogen in the purge gas. The remaining gas is recirculated to reactors DC-101B and DC-101C.

- The water phase from separator FA-104 goes to the PHOSAM Unit in the SNG plant to recover the H2S and NH3.

- The hydrocarbon phase from separator FA-104 is preheated in exchanger EA-105 and is combined with the hot liquid from FA-103 and charged to the HDT Fractionator DA-101. Fractionator DA-101 produces $500^{o}F$+ product (which is sent to hydrocracking, area 200), JP-4 (which is sent to storage), and unstabilized naphtha (which is sent to the naphtha stabilizer in the hydrocracking area 200).

- Catalyst is replaced every third day in each reactor so that one reactor is receiving and withdrawing catalyst each day. Catalyst is added and replaced by the catalyst handling system.

- Waste heat is converted to 100 psig saturated steam in exchangers EA-107 and EA-108. This steam is used for stripping in DA-101 and in the hydrocracking area 200 for stripping steam. There is an excess of about 5280 #/hr which is exported to the SNG steam system.

Table 2.1 Hydrotreater Conditions

Case 1 Maximum JP-4 Operation

Reactor Type	Expanded Bed
Number of Reactors	3
Reactor Temperature	$700^{o}F$
Temp. rise/stage	$225^{o}F$ max.
Ratio of H2 in Feed to Chemical H2	2.0 min.
Catalyst Replacement	0.18 #/Bbl

2.2 Hydrocracker (Area 200)

Operating conditions for the hydrocracker were provided to Lummus by AMOCO and these conditions are presented in Table 2.2. The basic processing step selected was a 5 bed fixed bed reactor system with recycle of unconverted 500°F+ material. Beds 1 and 2 use a catalyst that is most active for sulfur, nitrogen and oxygen removal while beds 3,4,5 use a catalyst that is most active for hydrocracking. Referring to drawing D5571-10201 and the material balance printouts (Appendix B) the flow is as follows:

- Hydrotreated 500°F+ material from the hydrotreater (Area 100) enters the system from day tank FA-201 through feed pump GA-201 and is preheated in exchanger EA-201. The preheated feed is combined with unconverted bottoms from fractionator DA-201 (approximately 35% of the feed is recycled). The combined oils are then mixed with hot hydrogen coming from heater BA-201 and charged to the reactor.

  The combined feed to the first bed in the reactor is 670°F.

- In the first bed the temperature rises to about 685°F. Quench hydrogen is added to cool the effluent from the first bed to about 652°F. In each of the remaining beds quench hydrogen is added to cool the beds. The inlet and outlet temperatures from each bed are as follows:

	Inlet	Outlet
Bed 1	675	685
Bed 2	652	683
Bed 3	653	682
Bed 4	653	688
Bed 5	656	696

- The reactor effluent is cooled in EA-201 and passes into the high temperature/high pressure separator FA-202. Vapors from FA-202 are cooled in EA-202 and then air condenser EC-201. Water is injected into the condenser EC-201 to dissolve H_2S and NH_3 into a NH_4OH/NH_4HS solution. This solution is sent to the PHOSAM unit in the SNG plant.

- The cooled vapors pass into separator FA-203 and the overhead hydrogen rich gas is divided with the major portion being used as recycle gas to the reactors via compressor GB-202 and heater BA-201. A small portion of the gas is purged from the system as high pressure purge gas which goes to the Rectisol unit in the SNG plant.

2.2 Hydrocracker (Area 200)

- The hydrocarbon phase from separator FA-203 is heated in exchanger EA-202, combined with the hot oil from separator FA-202 and charged to fractionator DA-201.

- Fractionator DA-201 produces unstabilized naphtha (which is charged to naphtha stabilizer DA-203), JP4 (which is sent to product storage after cooling), 500^0F+ oil (which is recycled to reactor DC-201) and fuel gas (which flows to the boiler in the SNG plant).

- Unstabilized naphtha from DA-201 is combined with unstabilized naphtha from area 100 and charged to naphtha stabilizer DA-203 after being preheated in exchanger EA-205. Heat for reboiling the naphtha stabilizer is obtained by heat exchange with the hot jet fuel product.

- Fuel gas from the naphtha stabilizer is combined with fuel gas from FA-206 and is routed to the SNG boilers.

- The stabilized naphtha is cooled and sent to storage. It is also sent to the SNG boilers to be used as fuel.

Table 2.2 Hydrocracker Conditions

Reactor Conditions	5 Fixed Beds
Catalyst, % of Total & Type	
Bed 1	10%, HDS/HDN/HC
Bed 2	22.5%, HDS/HDN/HC
Beds 3-5	22.5% HC
WHSV, hr^{-1}	1.1
Average Reactor Temp.	670^0F
Temperature Increase	
Bed 1	25^0F
Bed 2-5	50^0F
Heat of Reaction	20,000 BTU/#Mole H2
Reactor Pressure	
Inlet	1200 psig
Outlet	1175 psig
Recycle Rate H2	15,000 scf/Bbl
Conversion/Pass	65%
Catalyst Replacement	3 years @ $6/#

2.3 PSA Hydrogen Unit & Recompression (Area 300)

2.3.1 Hydrogen for both the hydrotreator and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 ^oF
Composition	mol%
H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

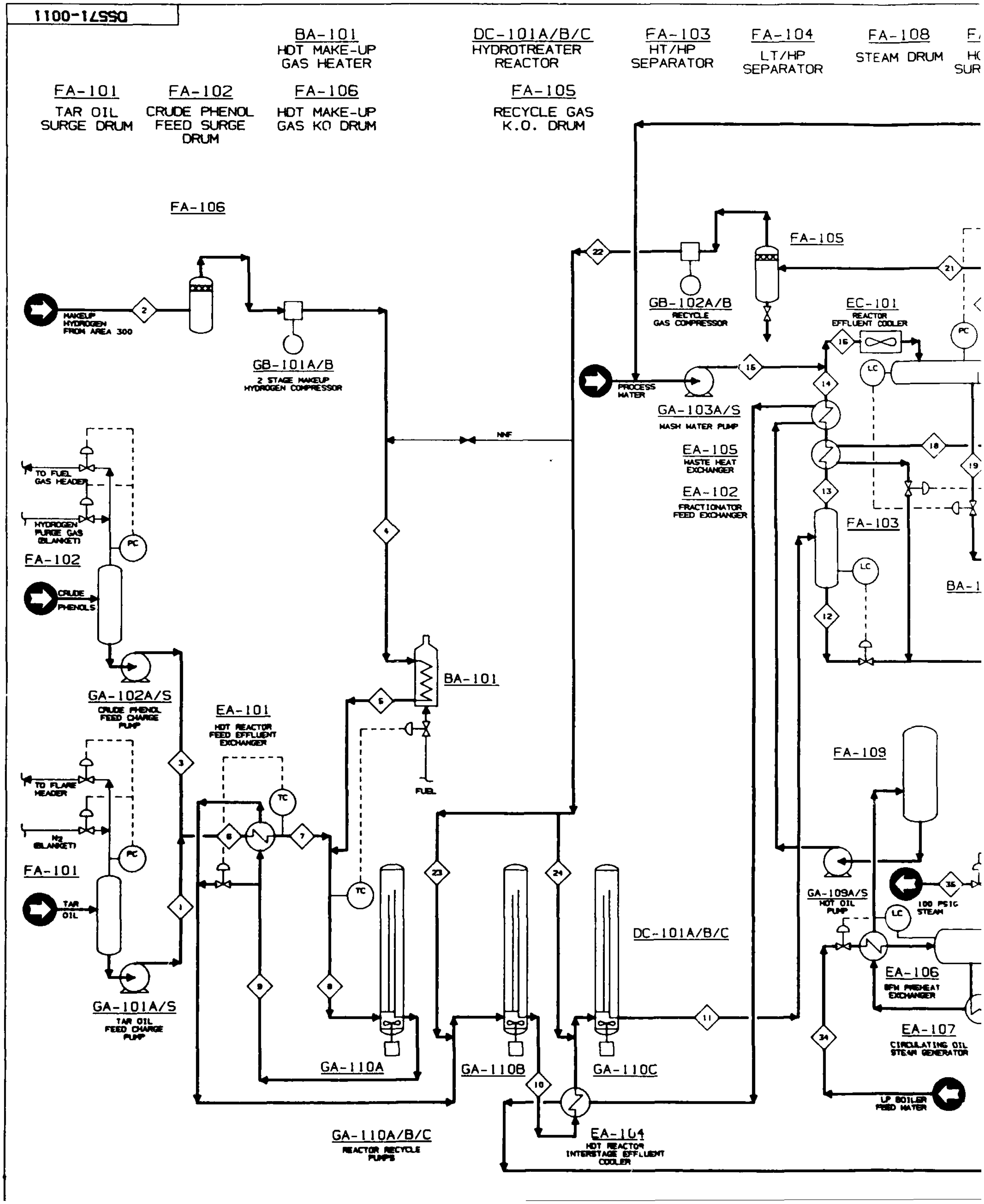
The PSA unit selectively absorbs all components expect H2 and produces a 99.99% vol. purity stream at about 345 psig and 80^oF. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100^oF
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

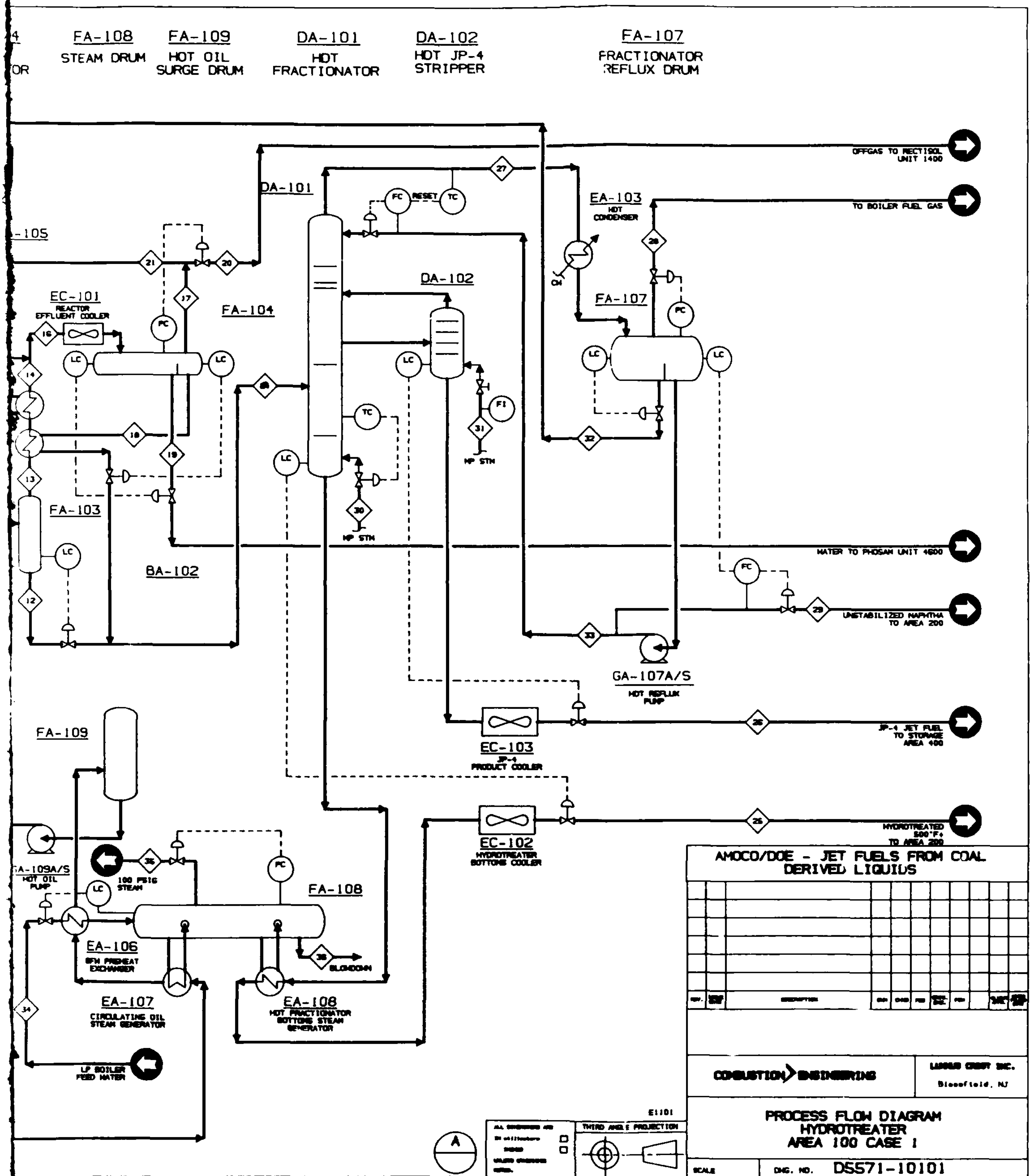
At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

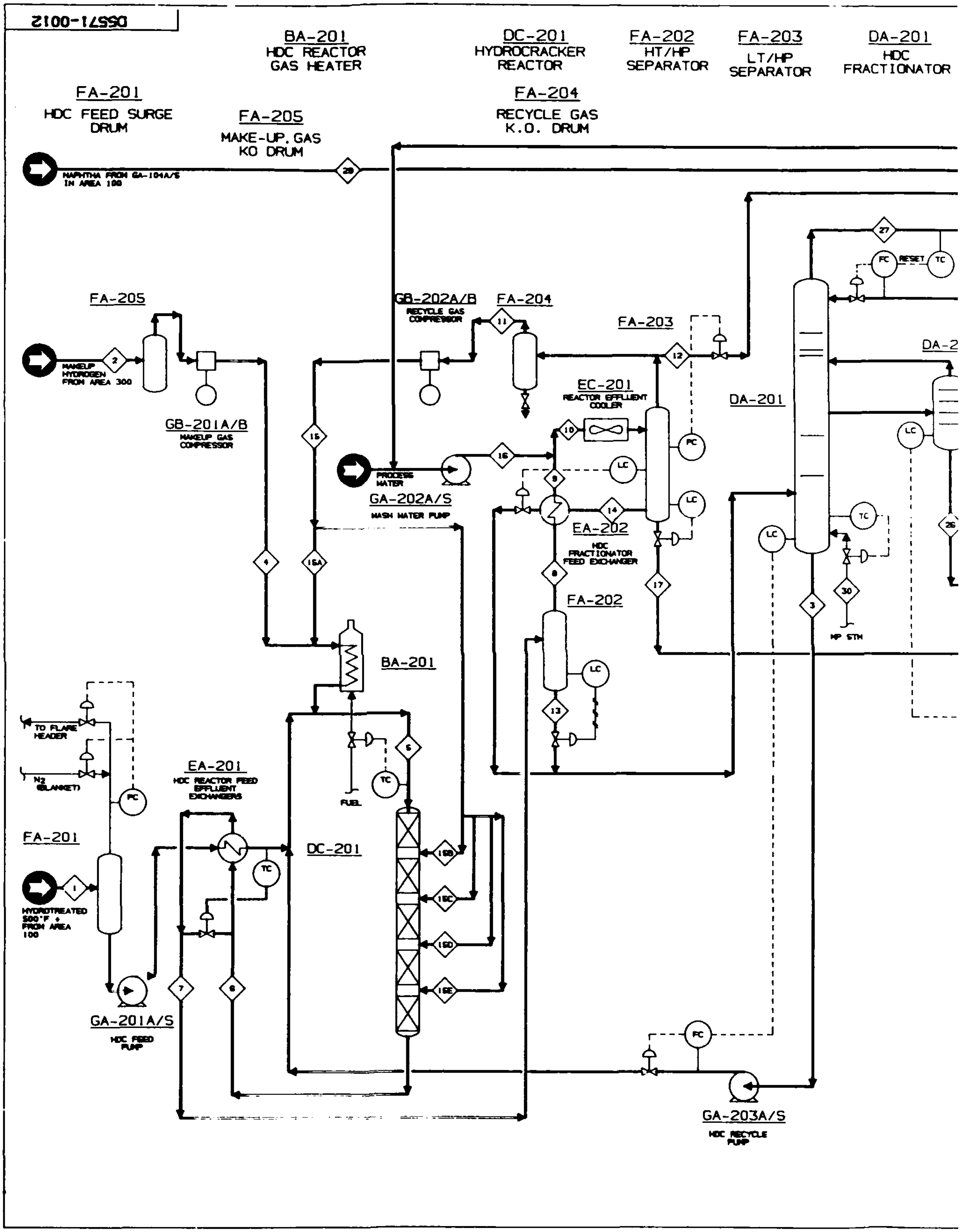
The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-10301 presents a schematic of a Union Carbide Polybed PSA unit.

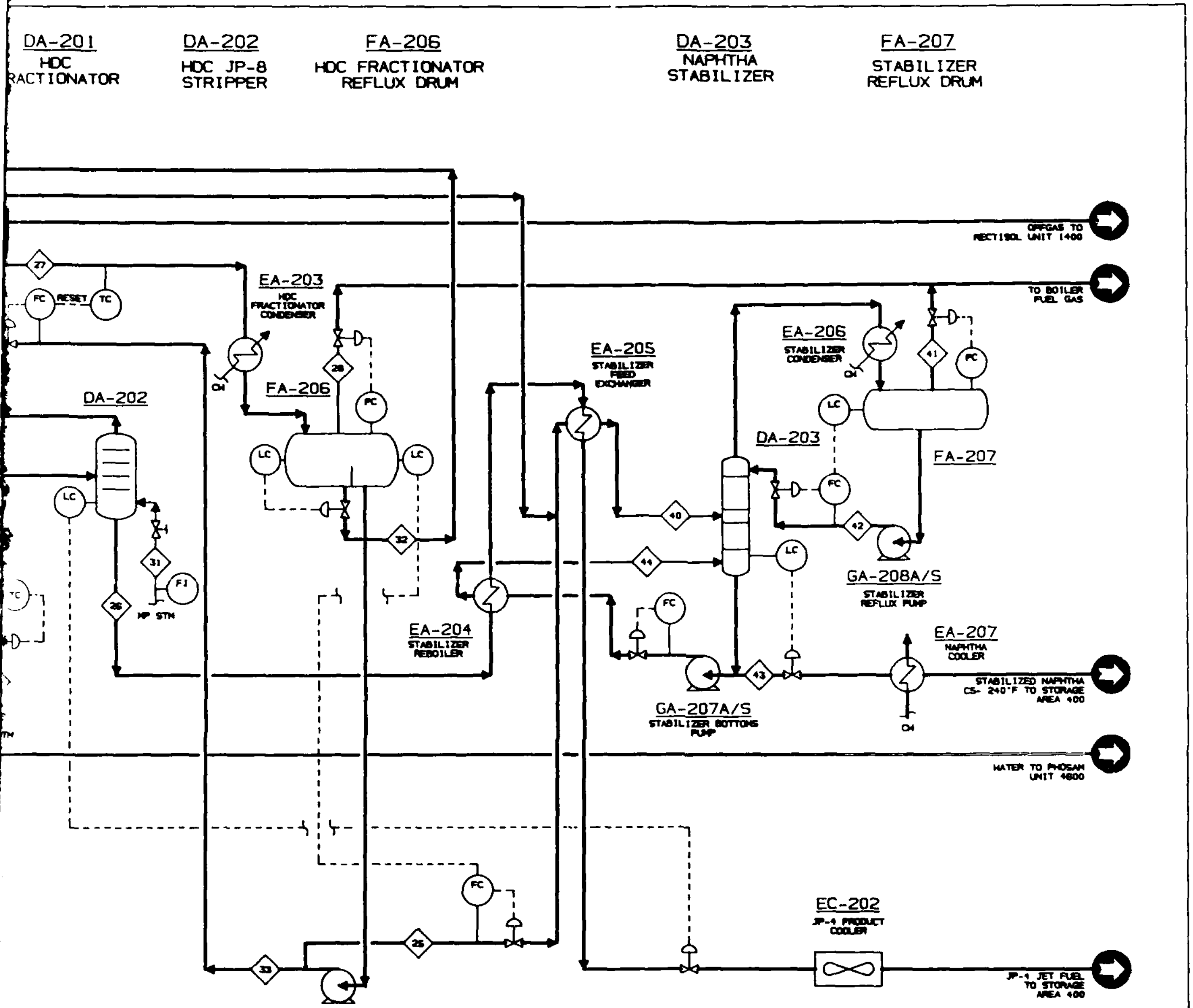
2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.



BA-101
HDT MAKE-UP GAS HEATER
DC-101A/B/C
HYDROTREATER REACTOR
FA-103
HT/HP SEPARATOR
FA-104
LT/HP SEPARATOR
FA-108
STEAM DRUM
FA-101
TAR OIL SURGE DRUM
FA-102
CRUDE PHENOL FEED SURGE DRUM
FA-106
HDT MAKE-UP GAS KO DRUM
FA-105
RECYCLE GAS K.O. DRUM
FA-106
MAKEUP HYDROGEN FROM AREA 300
GB-101A/B
2 STAGE MAKEUP HYDROGEN COMPRESSOR
FA-105
GB-102A/B
RECYCLE GAS COMPRESSOR
EC-101
REACTOR EFFLUENT COOLER
PROCESS WATER
GA-103A/S
WASH WATER PUMP
EA-105
WASTE HEAT EXCHANGER
EA-102
FRACTIONATOR FEED EXCHANGER
FA-103
NNF
TO FUEL GAS HEADER
HYDROGEN PURGE GAS (BLANKET)
FA-102
CRUDE PHENOLS
GA-102A/S
CRUDE PHENOL FEED CHARGE PUMP
EA-101
HDT REACTOR FEED EFFLUENT EXCHANGER
BA-101
FUEL
TO FLARE HEADER
N2 (BLANKET)
FA-101
TAR OIL
GA-101A/S
TAR OIL FEED CHARGE PUMP
FA-109
GA-109A/S
HOT OIL PUMP
100 PSIG STEAM
DC-101A/B/C
EA-106
BFW PREHEAT EXCHANGER
EA-107
CIRCULATING OIL STEAM GENERATOR
GA-110A
GA-110B
GA-110C
LP BOILER FEED WATER
GA-110A/B/C
REACTOR RECYCLE PUMPS
EA-104
HDT REACTOR INTERSTAGE EFFLUENT COOLER


FA-108
STEAM DRUM
FA-109
HOT OIL
SURGE DRUM
DA-101
HDT
FRACTIONATOR
DA-102
HDT JP-4
STRIPPER
FA-107
FRACTIONATOR
REFLUX DRUM
OFFGAS TO RECTISOL
UNIT 1400
TO BOILER FUEL GAS
EA-103
HDT
CONDENSER
DA-101
DA-102
FA-107
EC-101
REACTOR
EFFLUENT COOLER
FA-104
FA-103
BA-102
HP STM
WATER TO PHOSAM UNIT 4600
UNSTABILIZED NAPHTHA
TO AREA 200
GA-107A/S
HDT REFLUX
PUMP
FA-109
EC-103
JP-4
PRODUCT COOLER
JP-4 JET FUEL
TO STORAGE
AREA 400
EC-102
HYDROTREATER
BOTTOMS COOLER
HYDROTREATED
500°F+
TO AREA 200
100 PSIG
STEAM
FA-108
EA-106
BFW PREHEAT
EXCHANGER
BLOWDOWN
EA-107
CIRCULATING OIL
STEAM GENERATOR
EA-108
HDT FRACTIONATOR
BOTTOMS STEAM
GENERATOR
LP BOILER
FEED WATER
AMOCO/DOE - JET FUELS FROM COAL
DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
PROCESS FLOW DIAGRAM
HYDROTREATER
AREA 100 CASE 1
THIRD ANGLE PROJECTION
E1101
SCALE
DWG. NO.
DS571-10101


DSS71-0012
BA-201
HDC REACTOR GAS HEATER
DC-201
HYDROCRACKER REACTOR
FA-202
HT/HP SEPARATOR
FA-203
LT/HP SEPARATOR
DA-201
HDC FRACTIONATOR
FA-201
HDC FEED SURGE DRUM
FA-205
MAKE-UP GAS KO DRUM
FA-204
RECYCLE GAS K.O. DRUM
NAPHTHA FROM GA-104A/S IN AREA 100
MAKEUP HYDROGEN FROM AREA 300
GB-201A/B
MAKEUP GAS COMPRESSOR
GB-202A/B
RECYCLE GAS COMPRESSOR
EC-201
REACTOR EFFLUENT COOLER
PROCESS WATER
GA-202A/S
WASH WATER PUMP
EA-202
HDC FRACTIONATOR FEED EXCHANGER
RESET
HP STM
TO FLARE HEADER
N_2 (BLANKET)
EA-201
HDC REACTOR FEED EFFLUENT EXCHANGERS
FUEL
HYDROTREATED 500°F + FROM AREA 100
GA-201A/S
HDC FEED PUMP
GA-203A/S
HDC RECYCLE PUMP



AMOCO/DOE - JET FUELS FROM COAL DERIVED LIQUIDS

COMBUSTION ENGINEERING

LUMMUS CREST INC.
Bloomfield, NJ

E1084

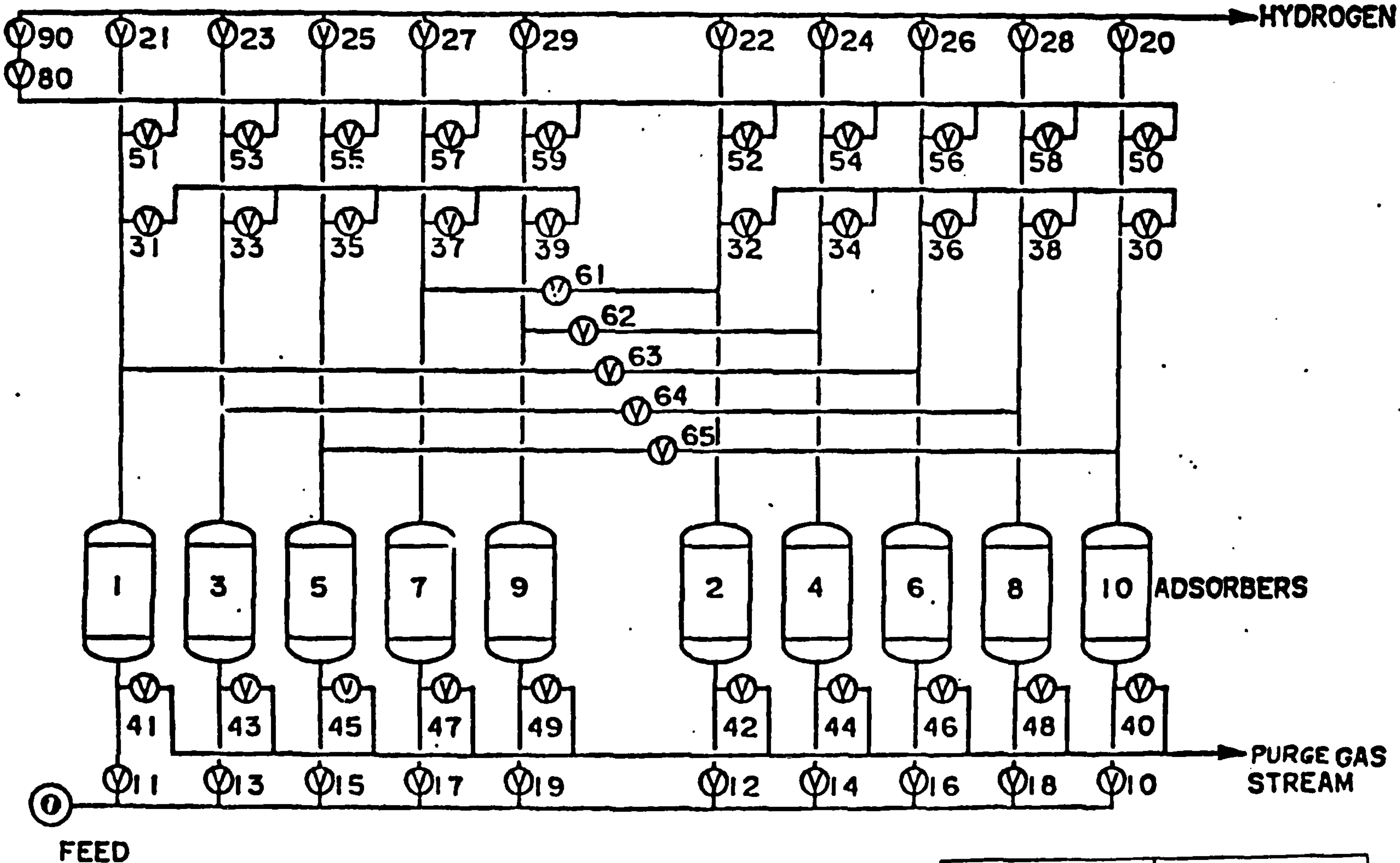
THIRD ANGLE PROJECTION

PROCESS FLOW DIAGRAM
500°F+ HYDROCRACKER
AREA 200 CASE 1

SCALE

DWG. NO. D5571-10201

SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 10

LUMMUS — THE LUMMUS COMPANY

TITLE PSA HYDROGEN UNIT AREA 300

CLIENT AMOCO/DOE

PROJ NO 5571

CASE 1

△				
△	1/3	FOR SUBTASK 1.2	ML	E.S.
			PROC	PROJ

5571- 3

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 1 - MAXIMUM JP-4

AREA 100 - HYDROTREATER

TAG NO. -	DESCRIPTION
BA-101	HDT Makeup Gas Heater
DA-101	HDT Fractionator
DA-102	JP-4 Stripper
DC-101A,B,C	Hydrotreater Reactors
EA-101	HDT Reactor Feed/Effl. Exch.
EA-102	Fract. Feed Exch.
EA-103	HDT Condenser
EA-104	HDT Reactor Int. Stg. Clr.
EA-105	Waste Heat Exchanger
EA-106	BFW Preheat Exch.
EA-107	Circulating Oil Stm. Gen.
EA-108	HTD Fract. Btms. Stm. Gen.
EC-101	Reactor Effl. Cooler
EC-102	Hydrotreater Btms. Cooler
EC-103	JP-4 Product Cooler
FA-101	Tar Oil Surge Drum
FA-102	Crude Phenol Surge Drum
FA-103	HT/HP Separator
FA-104	LT/HP Separator
FA-105	Recycle Gas KO Drum
FA-106	HDT Makeup Gas Ko Drum
FA-107	Fractionator Reflux Drum
FA-108	Steam Drum
FA-109	Hot Oil Surge Drum
GA-101A/S	Tar Oil Feed Charge Pump
GA-102A/S	Crude Phenol Feed Charge Pump
GA-103A/S	Wash Water Pump
GA-107A/S	HDT Reflux Pump
GA-109A/S	Hot Oil Pump
GA-110A/B/C	Reactor Recycle Pump
GB-101A/B	H_2 Makeup Compr.
GB-102A/B	Recycle Gas Compr.

CASE 1 - MAXIMUM JP-4 - Cont'd

AREA 200 - HYDROCRACKER

TAG. NO.	DESCRIPTION
BA-201	HDC Reactor Gas Heater
DA-201	HDC Fractionator
DA-202	HDC JP-4 Stripper
DA-203	Naphtha Stabilizer
DC-201	HDC Reactor
EA-201	HDC Reactor Feed/Effl. Exch.
EA-202	HDC Fract. Feed Exch.
EA-203	HDC Fract. Condenser
EA-204	Stabilizer Reboiler
EA-205	Stabilizer Feed Exch.
EA-206	Stabilizer Condenser
EA-207	Naphtha Cooler
EC-201	Reactor Effl. Cooler
EC-202	JP-4 Product Cooler
FA-201	HDC Feed Surge Drum
FA-202	HT/HP Separator
FA-203	LT/HP Separator
FA-204	Recycle Gas KO Drum
FA-205	Makeup Gas KO Drum
FA-206	HDC Fract. Reflux Drum
FA-207	Stabilizer Reflux Drum
FA-208	Fuel Oil Day Tank
GA-201A/S	HDC Feed Pump
GA-202A/S	Wash Water Pump
GA-203A/S	HDC Recycle Pump
GA-204A/S	HDC Naphtha Pump
GA-207A/S	Stabilizer Btms Pump
GA-208A/S	Stabilizer Reflux Pump
GA-209A/S	Fuel Oil Pump
GB-201A/B	Makeup Gas Compr.
GB-202A/B	Recycle Gas Compr.

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

CASE 1 - MAXIMUM JP-4 - Cont'd

TAG NO.	DESCRIPTION
AREA 400 -	STORAGE AREA
FB-401	Jet Fuel Storage Tank
FB-402	Naphtha Storage Tank
FB-403	Fuel Oil Storage Tank
GA-401A/S	Tar/Tar Oil Feed Pump
GA-402A/S	Crude Phenol Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump

AREA 500 - CATALYST HANDLING

TAG NO.	DESCRIPTION
FA-501	Catalyst Oil Drum
FA-502	Catalyst Storage Hopper
FA-503	Catalyst Transfer Vessels
FA-504	Spent Catalyst Vessel
FL-501	Catalyst Screen
GA-501A/S	Catalyst Transfer Pump
GA-502A/S	Catalyst Oil Pump

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for high pressure equipment purchased for various hydrotreater/hydrocracker projects. The unit pricing is some what conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the high pressure processing, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

	(Thousands of $) Case 1
Area 100 Hydrotreater	$23,836
Area 200 Hydrocracker	11,598
Area 300 PSA & Recompression	9,600
Area 400 OSBL	5,111
Area 500 Catalyst Handing	1,285
Total	$51,430

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

Items	Equipment Type	$ Value	%. Comm	$ Comm.
1	Heaters	80	120	96
2	Towers	55	140	77
	Internals	7	-	-
3	Reactors	2025	65	1316
12	Exchangers	625	70	438
3	Air Coolers	99	100	99
9	Vessels	429	85	365
13	Pumps	881	80	705
4	Compressors	2050	60	1230
	Special	-		
47	Total	$6251		$4326

Equipment	6251	
Commodities	4326	
Labor	3221	10% Equip. 60% Comm.
Indirects	3221	100%
Office	2844	47 pcs x 1100 x $55-
Subtotal	19,863	
Contingency	3,973	20%
Total	$23,836	

W16213/bf

3.2.3 Estimate Breakdown - Cont'd

Area 200

Items	Equipment Type	$ Valve	%. Comm	$ Comm.
1	Heaters	175	120	210
3	Towers	52	140	73
	Internals	8	-	-
1	Reactors	400	85	340
8	Exchangers	181	100	181
2	Air Coolers	57	110	63
8	Vessels	261	100	261
14	Pumps	162	120	194
4	Compressors	800	100	800
	Special	-		
41	Total	$2096		$2122

Equipment	2096	
Commodities	2122	
Labor	1483	10% Equip. 60% Comm.
Indirects	1483	100%
Office	2481	41 pcs x 1100 x $55-
Subtotal	9665	
Contingency	1933	20%
Total	$11,598	

Area 300

PSA-unit 20 mm SCFD budget quote $3500

PSA unit 17 mm SCFD 3000

Installation 50% 1500

Subtotal $4500

Compressor	3900	1950 x 2.0 T.I.C.
Drum	300	100 x 3.0 T.I.C.

Subtotal $8700

Contingency 900 10%

Total $9600

3.2.3 Estimate Breakdown - Cont'd

Area 400

Equipment & Value

Tankage		MTLS/C	805
Pumps		MTL	44
Yard Piping	28,400 LF	MTL	$ 475
Labor 22,000 Hrs x $50/hr		Labor S/C	$1,100
Pipe Insulation		S/C	160
Tracing		S/C	210
Excavation for U/G Pipe	55,000 Y^3	S/C	275
Rack	150 Ton Steel	S/C	300
Rack	FDN-500 Y^3	S/C	200

Equipment Related Commodities

Insulation 18000 Ft^2 x $10	S/C	180
FDNS, Instr, Elec, Paint - 20% x Equip.	MTL	170
S/C Labor @ 100% MTL	S/C	170
Total		$4,089
Contingency 25%		$1,022
Total		$5,111

Area 500

Items	Equipment Type	$ Value	%. Comm	$ Comm.
4	Vessels	105	120	126
4	Pumps	48	120	58
8	Total	$153		$184

Equipment	153	
Commodities	184	
Labor	125	10% Equip. 60% Comm.
Indirects	125	100%
Office	484	8 pcs x 1100 x $55-
Subtotal	$1071	
Contingency	214	20%
	$1285	

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require 7 men/shift to operate the plant broken down as follows:

Foreman	1
Control Room	1
HDT Operator	2
HCR Operator	2
PSA & relief man	1
	7 Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 5 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	7 positions x 4 people/position -	28
Supervisor & Admin.		5
QC Technician		1
Maintenance		5
Other (Stores or Janitorial)		1
Total		40

4.2 Utilities

The following utilities have been estimated from the computer simulations:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	3372 BPSD	$16/Bbl [a]	53952
SNG equivalent of Syn Gas & Purge Gas	4.81 MMSCFD	$3.80/MM Btu [b]	17912
Cooling Water	2400 GPM	$0.155/MGal [c]	536
Power	7100 kW	$0.04/kWH [c]	6816
Process Water	18.5 GPM	$0.45/MGal [c]	12

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

Catalyst	Use	Cost	$/SD
HDT Cat.	0.18 #/Bbl	$3.00/#	2218
HCR Cat.	0.013 #/Bbl	$6.00/#	96
Inhibitors	50 PPM	$10/Gal	86
			2400

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

$$0.00005 \times 51{,}430{,}000 = \$2571/SD$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-4 are proposed to be located to the east of the Rectisol Unit and Main Control Room of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 300' x 220' will be surrounded by an access road and will be divided by a central east-west road. Areas 100 & 500 will be located to the north and Areas 200 & 300 to the south.

A diked storage tank area approx. 360' x 265' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

Approximately 2000 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

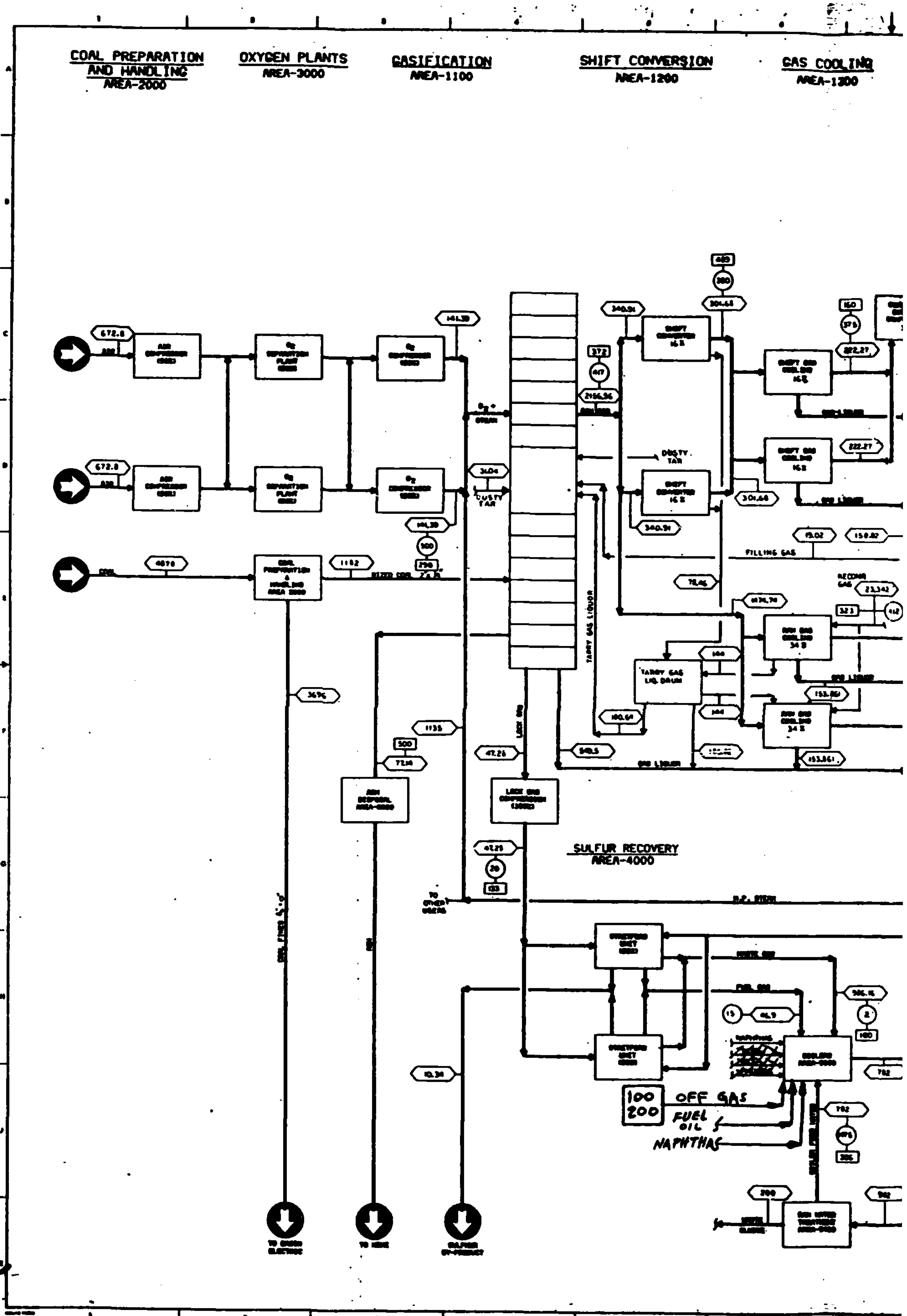
A summary of the lines is shown in table 5.1.

TABLE 5.1

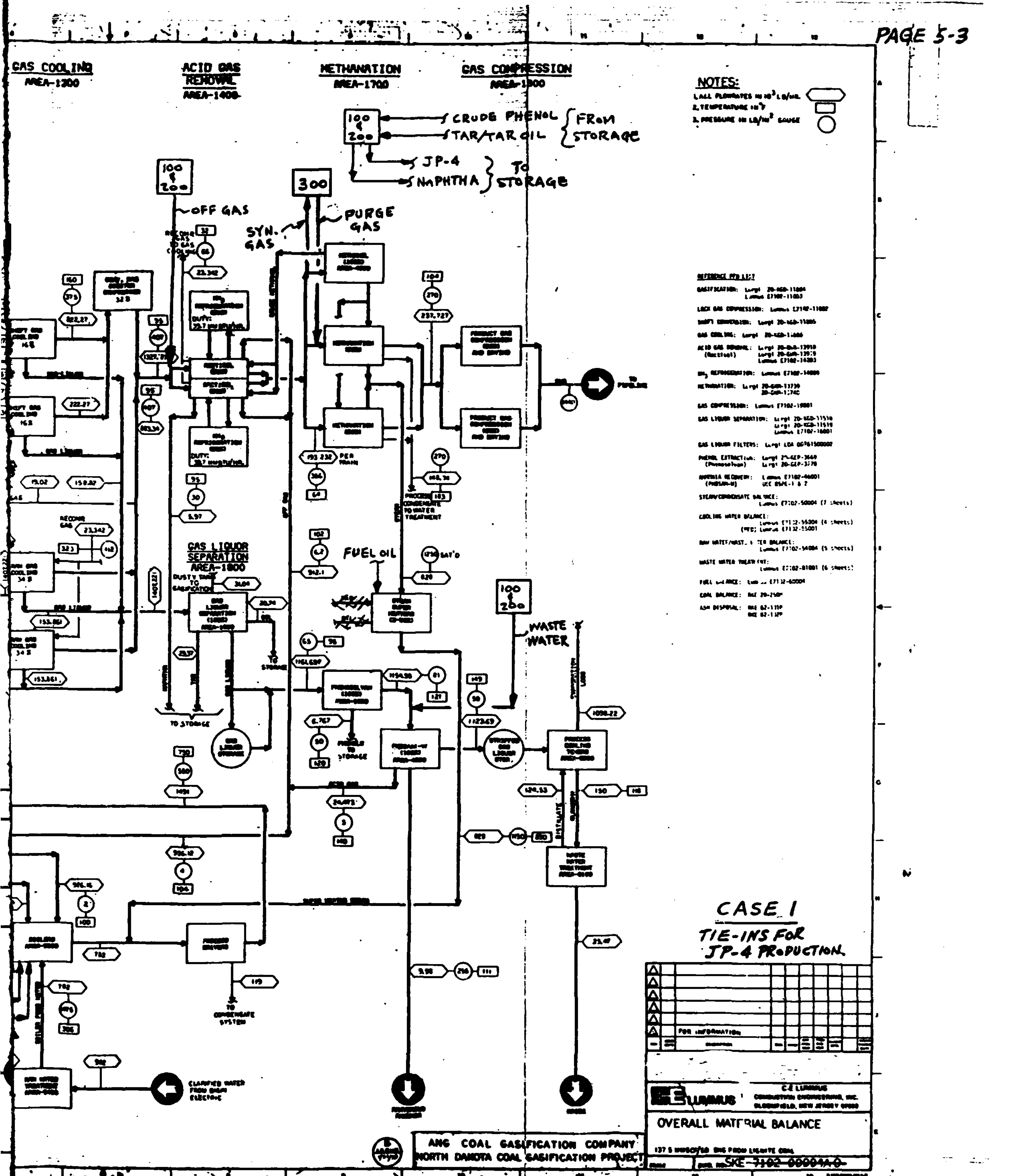
INTERCONNECTING PIPING

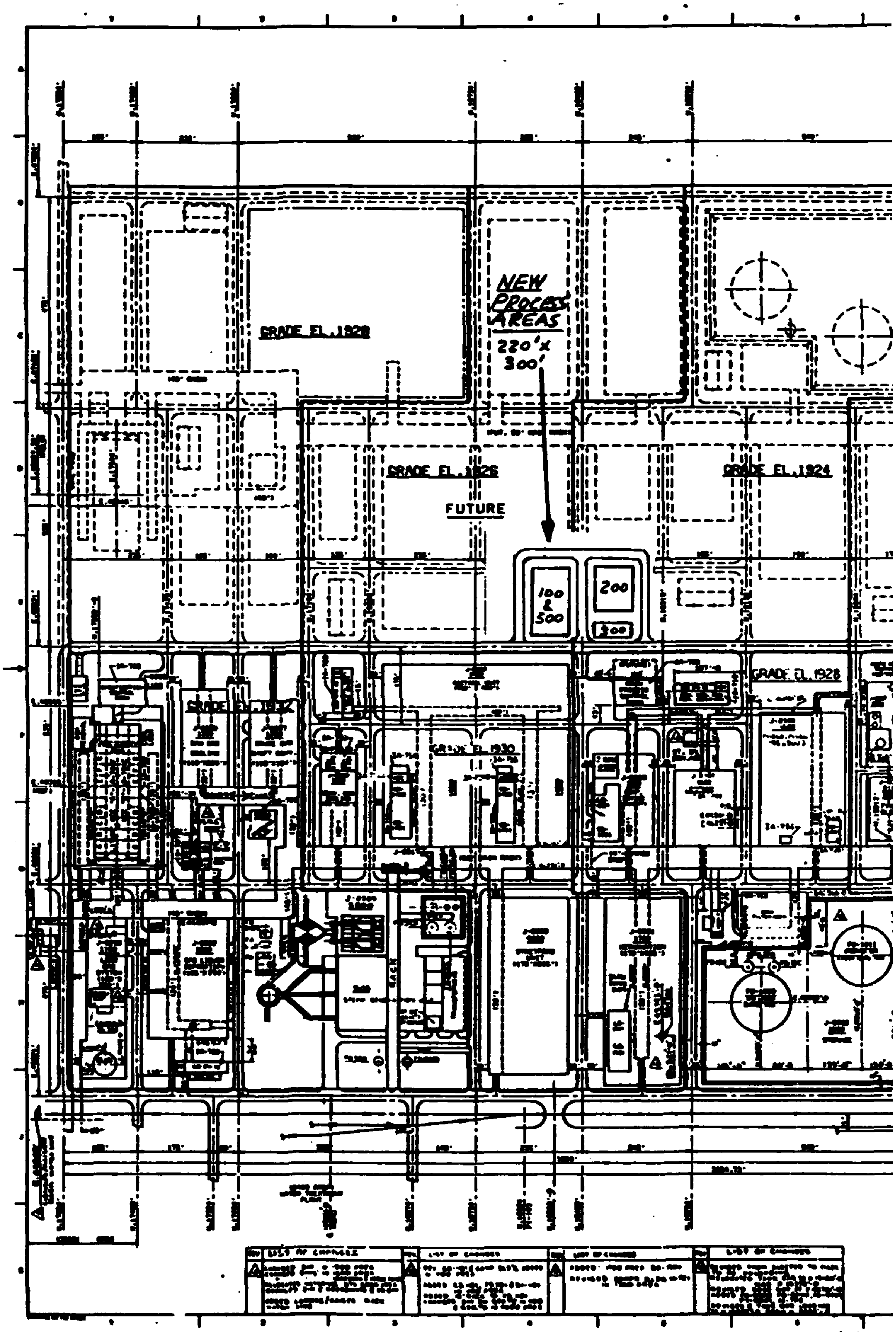
EST. SIZE	SERVICE	TO/FROM
4"	Tar/Tar Oil (Elec. Tr.)	Storage
2"	Crude Phenol (Elec. Tr.)	Storage
4"	JP-4 Product	Storage
1 1/2	Naphtha Product	Storage
18"	Wet Flare (Trace)	Flare
8"	Synthesis Gas	PSA/Rectisol
6"	Purge Gas	Methanation/PSA
2"	Off Gas	Rectisol/HDT,HDC
2"	Nitrogen	Main Rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
6"	M.P. Steam	"
1 1/2	Stm Cond.	"
1 1/2	BFW	"
1 1/2	Boiler B.D.	"
12"	C. W. Supply & Return	"
2"	Waste Water	Phosam/HDT,HDC
6"	Fuel Oil	Exist TKS/New TKS.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers

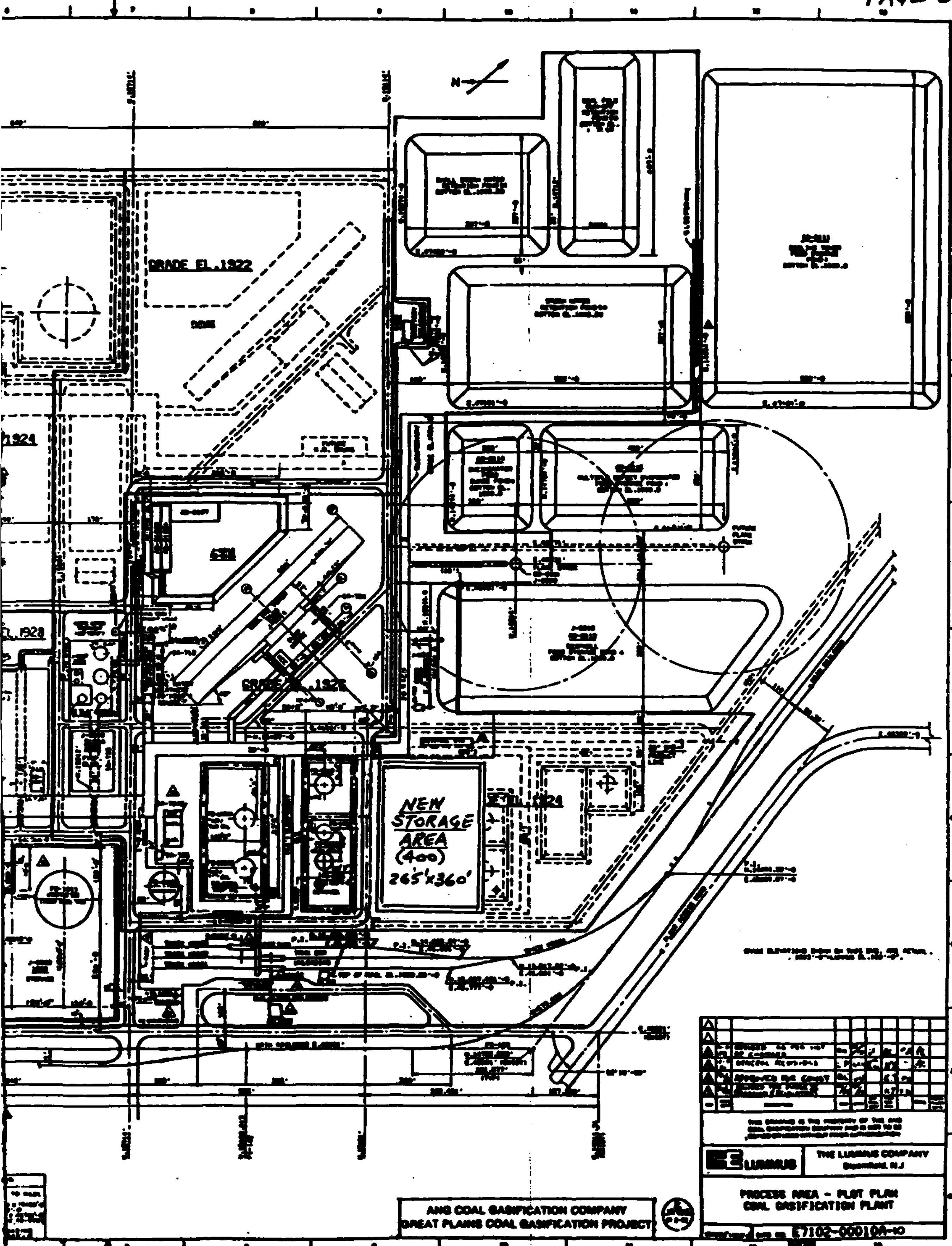
W16213/bf



COAL PREPARATION AND HANDLING AREA-2000
OXYGEN PLANTS AREA-3000
GASIFICATION AREA-1100
SHIFT CONVERSION AREA-1200
GAS COOLING AREA-1300
SULFUR RECOVERY AREA-4000
AIR
COAL
DUSTY TAR
TARRY GAS LIQUOR
FILLING GAS
LOCK GAS
OFF GAS
FUEL OIL
NAPHTHAS
100
200


GAS COOLING
AREA-1200
ACID GAS REMOVAL
AREA-1400
METHANATION
AREA-1700
GAS COMPRESSION
AREA-1900
NOTES:
100 & 200
CRUDE PHENOL
TAR/TAR OIL
FROM STORAGE
JP-4
NAPHTHA
TO STORAGE
300
OFF GAS
SYN. GAS
PURGE GAS
GAS LIQUOR SEPARATION
AREA-1800
FUEL OIL
WASTE WATER
TO STORAGE
CASE I
TIE-INS FOR JP-4 PRODUCTION
LUMMUS
OVERALL MATERIAL BALANCE
ANG COAL GASIFICATION COMPANY
NORTH DAKOTA COAL GASIFICATION PROJECT
SKE-7102-00004A-0


NEW PROCESS AREAS
220'x 300'
GRADE EL.1928
GRADE EL.1926
FUTURE
GRADE EL.1924
100 & 500
200
300
GRADE EL.1928
GRADE EL.1930



N
GRADE EL.1922
GRADE EL.1926
EL.1928
EL.1924
NEW STORAGE AREA (400) 265'x360'
ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT
LUMMUS
THE LUMMUS COMPANY
PROCESS AREA - PLOT PLAN
COAL GASIFICATION PLANT
E7102-00010A-10

CASE 1

VERSION 2.01
SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM C1U100

PROCESS TM
SOLUTION

HYDROTREATER

STREAM ID.	1	2	3	4	5	6
STREAM NAME..........	FEED TAR	MAKEUP H2	CP PHENOLS	NET MAKEUP	H2 FR B101	COMBINE FEED
STREAM PHASE.........	LIQUID	VAPOR	LIQUID	VAPOR	VAPOR	LIQUID
TEMPERATURE, DEG F	154.0000	70.0000	120.0000	393.0950	576.2288	147.7182
RATE LB MOLS/HR	304.2547	1800.0000	141.7299	1799.9998	1799.9998	446.0147
RATE LB /HR	47619.9688	3631.3228	14212.9863	3631.3228	3631.3228	61832.9141
ENTHALPY MM BTU /HR	2.0423	-3.7421	0.4788	0.1852	2.4813	2.5211
ENTHALPY BTU /LB	42.8877	-1030.5193	33.6883	51.0052	683.2996	40.7731
MOLECULAR WEIGHT....	156.4980	2.0174	100.2522	2.0174	2.0174	138.6342
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	3631.3228	0.0000	3631.3228	3631.3228	0.0000
STD.RATE MM FT3/DAY	0.00	16.39	0.00	16.39	16.39	0.00
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4475	3.4485	0.0000
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	2.0174	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	47619.9688	0.0000	14212.9863	0.0000	0.0000	61832.9141
ACT.RATE BBL/DAY	3261.67	0.00	929.35	0.00	0.00	4195.35
STD. LV RATE BBL/HR	132.85	0.00	38.20	0.00	0.00	171.06
CP, BTU /LB F	0.4043	0.0000	0.4025	0.0000	0.0000	0.4044
MOLECULAR WEIGHT....	156.4980	0.0000	100.2822	0.0000	0.0000	138.6342
ACT.DENS LB /FT3	62.4044	0.0000	65.3727	0.0000	0.0000	63.0005
STD. API GRAVITY....	6.5894	0.0000	1.5224	0.0000	0.0000	5.4247
*** DRY BASIS ***						
RATE LB /HR	46667.9766	0.0000	13573.0332	0.0000	0.0000	60240.9688
MOLECULAR WEIGHT....	185.6023	0.0000	127.7986	0.0000	0.0000	168.4372
UOP K	9.9520	0.0000	8.9460	0.0000	0.0000	9.7261
FLASH POINT, DEG F	142.3676	0.0000	92.6111	0.0000	0.0000	119.6567
CRIT. TEMP, F	966.3024	0.0000	786.3197	0.0000	0.0000	912.8544
CRIT. PRES, PSIA	424.5859	0.0000	546.5264	0.0000	0.0000	461.0103
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	3631.3228	0.0000	3631.3228	3631.3228	0.0000
STD.RATE MM FT3/HR	0.00	0.68	0.00	0.68	0.68	0.00
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4475	3.4485	0.0000
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	2.0174	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	46667.9766	0.0000	13573.0332	0.0000	0.0000	60240.9688
ACT.RATE BBL/DAY	3195.27	0.00	855.02	0.00	0.00	4084.23
CP, BTU /LB F	0.3923	0.0000	0.3749	0.0000	0.0000	0.3889
MOLECULAR WEIGHT....	185.6025	0.0000	127.7986	0.0000	0.0000	168.4372
ACT.DENS LB /FT3	62.4312	0.0000	65.5563	0.0000	0.0000	63.0484
STD. API GRAVITY....	[illegible]	[illegible]	1.1147	0.0000	0.0000	5.3021

VERSION 2.01
SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM C1U100

PROCESS™
SOLUTION

HYDROTREATER

STREAM ID.	7	8	9	10	11	12
STREAM NAME.........	HX FD EFF	101A INLET	101A OUTLET	101B OUTLET	101C OUTLET	HOT SEP LIQ
STREAM PHASE.........	LIQUID	MIXED	MIXED	MIXED	MIXED	LIQUID
TEMPERATURE, DEG F	503.5925	472.5212	700.0000	700.0000	700.0000	699.9121
RATE LB MOLS/HR	446.0145	2245.0112	1726.6343	1684.4119	2182.7847	85.1379
RATE LB /HR	61832.9141	65464.2422	65463.9531	66547.2969	69445.2969	10872.4727
ENTHALPY MM BTU /HR	13.1170	15.5983	30.0528	32.8555	38.0253	3.8500
ENTHALPY BTU /LB	212.1363	238.2719	459.0744	493.7171	547.5342	354.1057
MOLECULAR WEIGHT....	138.6343	29.1469	36.6000	39.5077	31.8164	127.7042
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	3441.2051	29373.8328	38375.8750	58528.3828	0.0000
STD.RATE MM FT3/DAY	0.00	16.88	13.86	13.30	19.10	0.00
CP, BTU /LB F	0.0000	1.6970	0.7879	0.7396	0.7558	0.0000
MOLECULAR WEIGHT....	0.0000	4.5550	19.2980	26.2875	27.9085	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	61832.9141	57023.0391	36090.0703	28171.4063	10919.9219	10872.4727
ACT.RATE BBL/DAY	4361.16	4640.74	3652.77	2990.44	1189.07	1184.28
STD. LV RATE BBL/HR	171.06	161.36	107.76	98.32	35.61	35.46
CP, BTU /LB F	0.5751	0.5505	0.6694	0.7192	0.7582	0.7579
MOLECULAR WEIGHT....	138.6343	145.1591	135.4171	125.4531	127.5177	127.7042
ACT.DENS LB /FT3	54.3716	52.5237	42.2335	40.2685	39.2558	39.2436
STD. API GRAVITY....	5.4247	8.5542	16.2850	23.6696	29.8915	29.8532
*** DRY BASIS ***						
RATE LB /HR	60240.9688	56767.1094	35702.9438	27655.0273	10731.0762	10685.0938
MOLECULAR WEIGHT....	165.4372	149.9299	145.7165	141.1739	142.7919	142.9699
UOP K	9.7261	9.8267	10.2764	10.6338	10.9789	10.9781
FLASH POINT, DEG F	119.6567	25.3483	-19.0451	-50.2837	-54.2565	-54.1564
CRIT. TEMP, F	912.8544	751.4496	644.9381	605.8790	569.6893	570.5956
CRIT. PRES, PSIA	461.0103	419.8909	373.5549	374.0819	356.0148	355.9686
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	7105.1943	26169.4531	33370.1328	51425.2969	0.0000
STD.RATE MM FT3/HR	0.00	0.58	0.51	0.45	0.65	0.00
CP, BTU /LB F	0.0000	1.9196	0.8222	0.7718	0.7885	0.0000
MOLECULAR WEIGHT....	0.0000	3.9939	19.4677	28.2321	30.1993	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	60240.9688	56767.1094	35702.9438	27655.0273	10731.0762	10685.0938
ACT.RATE BBL/DAY	4721.36	4619.02	3591.31	2908.48	1159.10	1154.58
CP, BTU /LB F	0.5591	0.5476	0.6385	0.6669	0.7096	0.7096
MOLECULAR WEIGHT....	165.4372	149.9299	145.7165	141.1739	142.7919	142.9699
ACT.DENS LB /FT3	54.5346	52.5338	42.4955	40.6444	39.5746	39.5590
STD. API GRAVITY....	5.1031	8.5474	15.3525	23.9237	30.2404	30.2002
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

VERSION 2.01
SIMULATION SCIENCES INC. PROCESS™ PAGE 21
PROJECT GP JET FUELS MAX JP4
PROBLEM C1U100 SOLUTION OCT 16 1987

HYDROTREATER

STREAM ID.	13	14	15	16	17	18
STREAM NAME..........	HOT SEP VAP	COCL RX EFF	H2O WASH EFF	+ H2O	COLD SEP VAP	COLD SEP LIQ
STREAM PHASE.........	VAPOR	MIXED	LIQUID	MIXED	VAPOR	LIQUID
TEMPERATURE, DEG F	699.9121	437.5901			120.0000	120.0000
RATE LB MOLS/HR	2097.6475	2097.6465			1260.6277	422.5313
RATE LB /HR	58575.8281	58575.8281			4063.7197	47032.2188
ENTHALPY MM BTU /HR	34.1750	19.0475			-1.8977	0.5580
ENTHALPY BTU /LB	583.4320	325.1774			-466.9748	11.8645
MOLECULAR WEIGHT....	27.9245	27.9245			3.2236	111.3106
*** VAPOR PHASE ***						
RATE LB /HR	58575.8281	20842.4219			4063.7197	0.0000
STD.RATE MM FT3/DAY	19.10	15.64			11.48	0.00
CP, BTU /LB F	0.7557	0.9528			2.2613	0.0000
MOLECULAR WEIGHT....	27.9245	12.1399			3.2236	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	0.0000	37733.3906			0.0000	47032.2188
ACT.RATE BBL/DAY	0.00	3965.64			0.00	4049.25
STD. LV RATE BBL/HR	0.00	130.05			0.00	165.76
CP, BTU /LB F	0.0000	0.6447			0.0000	0.4640
MOLECULAR WEIGHT....	0.0000	99.0930			0.0000	111.3106
ACT.DENS LB /FT3	0.0000	40.6708			0.0000	49.6492
STD. API GRAVITY....	0.0000	39.0882			0.0000	42.9141
*** DRY BASIS ***						
RATE LB /HR	0.0000	36473.8750			0.0000	47017.0154
MOLECULAR WEIGHT....	0.0000	117.3274			0.0000	111.4973
UOP K	0.0000	11.1838			0.0000	11.2645
FLASH POINT, DEG F	0.0000	-61.5335			0.0000	-77.8154
CRIT. TEMP, F	0.0000	525.6302			0.0000	549.8193
CRIT. PRES, PSIA	0.0000	407.3491			0.0000	478.1943
*** VAPOR PHASE ***						
RATE LB /HR	51471.2313	14997.3945			4044.3599	0.0000
STD.RATE MM FT3/HR	0.65	0.53			0.48	0.00
CP, BTU /LB F	0.7883	1.0320			2.2688	0.0000
MOLECULAR WEIGHT....	30.2189	10.7709			3.2109	0.0000
VISCOSITY, CP	0.0000	0.0000			0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	0.0000	36473.8750			0.0000	47017.0154
ACT.RATE BBL/DAY	0.00	3562.36			0.00	4048.22
CP, BTU /LB F	0.0000	0.6263			0.0000	0.4638
MOLECULAR WEIGHT....	0.0000	117.3274			0.0000	111.2645
ACT.DENS LB /FT3	0.0000	40.3665			0.0000	49.6461
STD. API GRAVITY....	0.0000	40.0905			0.0000	42.9551

VERSION 2.01 SIMULATION SCIENCES INC. PROCESS™ SOLUTION PAGE 22
PROJECT GP JET FUELS MAX JP4
PROBLEM C1U1UU OCT 16 1987

HYDROTREATER

STREAM ID.	19	20	21	22	23	24
STREAM NAME..........	SOUR H2O	PURGE GAS	RECYCLE GAS	REC COMP DIS	101B REC GAS	101C REC GAS
STREAM PHASE.........	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F		120.0000	120.0000	138.5653	138.5652	138.5652
RATE LB MOLS/HR		24.4908	1236.1367	1236.1384	336.1385	899.9999
RATE LB /HR		78.9478	3984.7710	3984.7793	1083.5659	2901.2119
ENTHALPY MM BTU /HR		-0.0369	-1.8608	-1.7003	-0.4624	-1.2380
ENTHALPY BTU /LB		-466.9752	-466.9749	-426.7032	-426.7034	-426.7035
MOLECULAR WEIGHT....		3.2236	3.2236	3.2236	3.2236	3.2236
*** VAPOR PHASE ***						
RATE LB /HR		78.9478	3984.7710	3984.7793	1083.5659	2901.2119
STD.RATE MM FT3/DAY		0.22	11.26	11.26	3.06	8.20
CP, BTU /LB F		2.2603	2.2603	2.2588	2.2588	2.2588
MOLECULAR WEIGHT....		3.2236	3.2236	3.2236	3.2236	3.2236
*** LIQUID PHASE ***						
RATE LB /HR		0.0000	0.0000	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY		0.00	0.00	0.00	0.00	0.00
STD. LV RATE BBL/HR		0.00	0.00	0.00	0.00	0.00
CP, BTU /LB F		0.0000	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....		0.0000	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3		0.0000	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....		0.0000	0.0000	0.0000	0.0000	0.0000
*** DRY BASIS ***						
RATE LB /HR		0.0000	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....		0.0000	0.0000	0.0000	0.0000	0.0000
UOP K		0.0000	0.0000	0.0000	0.0000	0.0000
FLASH POINT, DEG F		0.0000	0.0000	0.0000	0.0000	0.0000
CRIT. TEMP, F		0.0000	0.0000	0.0000	0.0000	0.0000
CRIT. PRES, PSIA		0.0000	0.0000	0.0000	0.0000	0.0000
*** VAPOR PHASE ***						
RATE LB /HR		78.5717	3965.7876	3965.7964	1078.4041	2887.3906
STD.RATE MM FT3/HR		0.01	0.47	0.47	0.13	0.34
CP, BTU /LB F		2.2639	2.2688	2.2673	2.2673	2.2673
MOLECULAR WEIGHT....		3.2109	3.2109	3.2110	3.2109	3.2109
VISCOSITY, CP		0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR		0.0000	0.0000	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY		0.00	0.00	0.00	0.00	0.00
CP, BTU /LB F		0.0000	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....		0.0000	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3		0.0000	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....		0.0000	0.0000	0.0000	0.0000	0.0000

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.		25	26	27	28	29
STREAM NAME.........	TOT FRAC FD	HDC FEED	HDT JP-4	FRAC OVHD	HDT HC VAP	HDT NAPHTHA
STREAM PHASE........	MIXED	LIQUID	LIQUID	VAPOR	VAPOR	LIQUID
TEMPERATURE, DEG F	549.6777	504.7608	329.8319	236.9774	90.0000	90.0000
PRESSURE, PSIA	64.0000	93.8995	95.5000	90.0000	80.0000	80.0000
RATE LB MOLS/HR	507.6690	79.5934	294.0670	1010.1720	91.2511	44.8024
RATE LB /HR	57904.6406	17056.0039	35947.9922	59035.8672	1727.7163	3209.6353
ENTHALPY MM BTU /HR	19.5373	3.1887	4.0676	15.8779	0.3145	0.0086
ENTHALPY BTU /LB	337.4048	186.9537	113.1529	268.9532	182.0228	2.6839
MOLECULAR WEIGHT....	114.0598	214.2392	122.2442	58.4414	18.9337	71.6398
*** VAPOR PHASE ***						
RATE LB /HR	49181.0078	0.0000	0.0000	59035.8672	1727.7163	0.0000
ACT.RATE FT3/SEC	20.87	0.00	0.00	21.61	1.84	0.00
STD.RATE MM FT3/DAY	4.25	0.00	0.00	9.20	0.83	0.00
CP, BTU /LB F	0.5800	0.0000	0.0000	0.4839	0.5648	0.0000
MOLECULAR WEIGHT....	105.5102	0.0000	0.0000	58.4414	18.9337	0.0000
ACT.DENS LB /FT3	0.6547	0.0000	0.0000	0.7590	0.2606	0.0000
COMPRESSIBILITY (Z).	0.9522	0.0000	0.0000	0.9270	0.9854	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	8723.6426	17056.0039	35947.9922	0.0000	0.0000	3209.6353
ACT.RATE BBL/DAY	816.86	1509.15	3536.36	0.00	0.00	318.73
STD. LV RATE BBL/HR	26.62	51.48	124.85	0.00	0.00	12.98
CP, BTU /LB F	0.6210	0.5924	0.5779	0.0000	0.0000	0.5048
MOLECULAR WEIGHT....	209.9675	214.2892	122.2442	0.0000	0.0000	71.6398
ACT.DENS LB /FT3	45.6503	48.3099	43.4521	0.0000	0.0000	43.0454
STD. API GRAVITY....	19.5506	17.8876	40.4044	0.0000	0.0000	68.6129
*** DRY BASIS ***						
RATE LB /HR	8723.1992	17007.3954	35772.2266	0.0000	0.0000	3208.8184
MOLECULAR WEIGHT....	210.1013	221.1758	125.8212	0.0000	0.0000	71.6941
UOP K	10.7600	10.7065	11.2069	0.0000	0.0000	11.7920
FLASH POINT, DEG F	110.4238	201.5718	48.4767	0.0000	0.0000	-74.4763
CRIT. TEMP, F	904.1510	946.4381	658.1058	0.0000	0.0000	422.6767
CRIT. PRES, PSIA	337.0012	322.0181	436.6072	0.0000	0.0000	575.3866
*** VAPOR PHASE ***						
RATE LB /HR	48978.8750	0.0000	0.0000	56144.7109	1713.4653	0.0000
ACT.RATE FT3/SEC	20.34	0.00	0.00	17.95	1.83	0.00
STD.RATE MM FT3/HR	0.17	0.00	0.00	0.32	0.03	0.00
CP, BTU /LB F	0.5807	0.0000	0.0000	0.4878	0.5658	0.0000
MOLECULAR WEIGHT....	107.6624	0.0000	0.0000	66.0780	18.9417	0.0000
ACT.DENS LB /FT3	0.6689	0.0000	0.0000	0.8688	0.2607	0.0000
COMPRESSIBILITY (Z).	0.9511	0.0000	0.0000	0.9156	0.9853	0.0000
VISCOSITY, CP	0.0127	0.0000	0.0000	0.0098	0.0110	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	8723.1992	17007.3954	35772.2266	0.0000	0.0000	3208.8184
ACT.RATE BBL/DAY	816.82	1504.89	3523.01	0.00	0.00	318.67
CP, BTU /LB F	0.6210	0.5206	0.5760	0.0000	0.0000	0.5047
MOLECULAR WEIGHT....	210.1013	221.1755	125.8212	0.0000	0.0000	71.6941
ACT.DENS LB /FT3	45.6503	45.3048	43.4035	0.0000	0.0000	43.0421
STD. API GRAVITY....	19.5511	17.9249	40.5534	0.0000	0.0000	68.6278
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]	0.0000	[illegible]

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	30	31	32	33
STREAM NAME..........	FRAC STR STM	JP-4 STR STM	SCUR H2O	REFLUX
STREAM PHASE.........	VAPOR	VAPOR	LIQUID	LIQUID
TEMPERATURE, DEG F	330.6417	331.3275	90.0000	90.0000
PRESSURE, PSIA	93.8996	95.5000	80.0000	80.0000
RATE LB MOLS/HR	72.1621	88.8149	158.9312	715.1871
RATE LB /HR	1299.9990	1599.9993	2863.1460	51235.8438
ENTHALPY MM BTU /HR	1.5473	1.9043	0.1661	0.1375
ENTHALPY BTU /LB	1190.2109	1190.2109	58.0156	2.6839
MOLECULAR WEIGHT....	18.0150	18.0150	18.0150	71.6398
*** VAPOR PHASE ***				
RATE LB /HR	1299.9990	1599.9993	0.0000	0.0000
ACT.RATE FT3/SEC	1.72	2.06	0.00	0.00
STD.RATE MM FT3/DAY	0.66	0.81	0.00	0.00
CP, BTU /LB F	0.5787	0.5805	0.0000	0.0000
MOLECULAR WEIGHT....	18.0150	18.0150	0.0000	0.0000
ACT.DENS LB /FT3	0.2100	0.2136	0.0000	0.0000
COMPRESSIBILITY (Z).	0.9493	0.9490	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	0.0000	0.0000	2863.1460	51235.8438
ACT.RATE BBL/DAY	0.00	0.00	197.08	5087.91
STD. LV RATE BBL/HR	0.00	0.00	8.19	207.15
CP, BTU /LB F	0.0000	0.0000	0.9978	0.5047
MOLECULAR WEIGHT....	0.0000	0.0000	18.0150	71.6398
ACT.DENS LB /FT3	0.0000	0.0000	62.1002	43.0454
STD. API GRAVITY....	0.0000	0.0000	10.0635	68.6129
*** DRY BASIS ***				
RATE LB /HR	0.0000	0.0000	0.0000	51222.8047
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	71.6941
UOP K	0.0000	0.0000	0.0000	11.7920
FLASH POINT, DEG F	0.0000	0.0000	0.0000	-74.4764
CRIT. TEMP, F	0.0000	0.0000	0.0000	422.6769
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	575.3867
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	0.0000	0.0000	0.0000	51222.8047
ACT.RATE BBL/DAY	0.00	0.00	0.00	5087.02
CP, BTU /LB F	0.0000	0.0000	0.0000	0.5046
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	71.6941
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	43.0420
STD. API GRAVITY....	0.0000	0.0000	0.0000	68.6278
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

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PROCESS(TM) SOLUTION

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CASE I
AREA 200

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	1	2	3	4	5	6
STREAM NAME.........	HDT SSC+	MAKEUP H2	HDC RECYCLE	NET MAKE-UP	RX INLET	RX EFFLUENT
STREAM PHASE.........	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	VAPOR
TEMPERATURE, DEG F	200.0000	70.0000	510.0000	443.2653	670.0000	696.1902
PRESSURE, PSIA	1240.0000	340.0000	1240.0000	1265.0000	1200.0000	1175.0000
RATE LB PCLS/HR	77.4065	250.0000	37.4100	250.0000	1461.0029	2367.8394
RATE LB /HR	17056.6367	504.3504	8901.9961	504.3504	30291.7891	33940.1719
ENTHALPY MM BTU /HR	0.4604	-0.5197	1.8367	0.1261	12.6494	18.7233
ENTHALPY BTU /LB	26.9926	-1030.5193	206.3200	249.9401	417.5854	551.6570
MOLECULAR WEIGHT....	220.3512	2.0174	237.9579	2.0174	20.7336	14.3338
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	504.3504	0.0000	504.3504	15859.0449	33940.1719
ACT.RATE FT3/SEC	0.00	1.18	0.00	0.55	4.03	7.17
STD.RATE MM FT3/DAY	0.00	2.29	0.00	2.28	12.65	21.57
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4758	1.0812	1.0240
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	11.4196	14.3338
ACT.DENS LB /FT3	0.0000	0.1189	0.0000	0.2543	1.0935	1.3156
COMPRESSIBILITY (Z).	0.0000	1.0145	0.0000	1.0357	1.0338	1.0321
*** LIQUID PHASE ***						
RATE LB /HR	17056.6367	0.0000	8901.9961	0.0000	14432.7441	0.0000
ACT.RATE BBL/DAY	1290.83	0.00	787.44	0.00	1501.74	0.00
STD. LV RATE BBL/HR	51.52	0.00	27.79	0.00	44.77	0.00
CP, BTU /LB F	0.4440	0.0000	0.6245	0.0000	0.6839	0.0000
MOLECULAR WEIGHT....	220.3512	0.0000	237.9579	0.0000	199.7616	0.0000
ACT.DENS LB /FT3	56.4828	0.0000	48.3241	0.0000	41.0815	0.0000
STD. API GRAVITY....	12.0000	0.0000	22.9965	0.0000	22.0322	0.0000
*** DRY BASIS ***						
RATE LB /HR	17056.6367	0.0000	8901.9961	0.0000	14432.1133	0.0000
MOLECULAR WEIGHT....	220.3513	0.0000	237.9579	0.0000	199.8498	0.0000
UOP K	10.7119	0.0000	11.1155	0.0000	10.9332	0.0000
FLASH POINT, DEG F	207.5695	0.0000	227.1025	0.0000	110.0334	0.0000
CRIT. TEMP, F	944.2889	0.0000	948.3727	0.0000	749.8058	0.0000
CRIT. PRES, PSIA	326.8333	0.0000	280.7643	0.0000	282.8885	0.0000
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	504.3504	0.0000	504.3504	15829.2266	33880.7188
ACT.RATE FT3/SEC	0.00	1.18	0.00	0.55	4.02	7.16
STD.RATE MM FT3/HR	0.00	0.09	0.00	0.09	0.53	0.90
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4758	1.0824	1.0249
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	11.4118	14.3287
ACT.DENS LB /FT3	0.0000	0.1169	0.0000	0.2543	1.0927	1.3151
COMPRESSIBILITY (Z).	0.0000	1.0145	0.0000	1.0357	1.0338	1.0322
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	17056.6367	0.0000	8901.9961	0.0000	14432.1133	0.0000
ACT.RATE BBL/DAY	1290.23	0.00	787.44	0.00	1501.66	0.00
CP, BTU /LB F	0.4440	0.0000	0.6245	0.0000	0.6838	0.0000
MOLECULAR WEIGHT....	220.3513	0.0000	237.9579	0.0000	199.8498	0.0000
ACT.DENS LB /FT3	56.4828	0.0000	48.3241	0.0000	41.0818	0.0000
STD. API GRAVITY....	[illegible]	[illegible]	22.9965	0.0000	22.0327	0.0000

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	7	3	9	11	12	13
STREAM NAME..........	E-202 OUTLET	HOT SEP VAP	EFF FR E-201	RECYCLE GAS	PURGE GAS	HOT SEP LIQ
STREAM PHASE.........	MIXED	VAPOR	MIXED	VAPOR	VAPOR	LIQUID
TEMPERATURE, DEG F	580.3386	580.2936	418.2935	120.0000	120.0000	580.2936
PRESSURE, PSIA	1150.0000	1145.0000	1120.0000	1095.0000	1095.0000	1145.0000
RATE LB MOLS/HR	2367.8403	2348.2407	2349.2417	2140.6616	10.4499	19.5989
RATE LB /HR	33940.1719	30240.9688	30240.9648	7476.9229	36.4994	3699.2319
ENTHALPY MM BTU /HR	14.4675	13.4662	7.6576	-3.0108	-0.0147	1.0021
ENTHALPY BTU /LB	426.2655	445.2978	253.2198	-402.6737	-402.6740	270.8992
MOLECULAR WEIGHT....	14.3338	12.8781	12.8781	3.4928	3.4928	188.7466
***** VAPOR PHASE *****						
RATE LB /HR	30215.8242	30240.9688	21046.9414	7476.9229	36.4994	0.0000
ACT.RATE FT3/SEC	6.53	6.56	5.53	3.52	0.02	0.00
STD.RATE MM FT3/DAY	21.39	21.39	20.86	19.50	0.10	0.00
CP, BTU /LB F	1.0318	1.0300	1.1587	2.1303	2.1303	0.0000
MOLECULAR WEIGHT....	12.8683	12.8781	9.1874	3.4928	3.4928	0.0000
ACT.DENS LB /FT3	1.2848	1.2804	1.0572	0.5895	0.5895	0.0000
COMPRESSIBILITY (Z).	1.0321	1.0319	1.0331	1.0430	1.0430	0.0000
***** LIQUID PHASE *****						
RATE LB /HR	3724.3750	0.0000	9194.0137	0.0000	0.0000	3699.2319
ACT.RATE BBL/DAY	387.13	0.00	915.60	0.00	0.00	384.37
STD. LV RATE BBL/HR	12.08	0.00	31.34	0.00	0.00	12.00
CP, BTU /LB F	0.6748	0.0000	0.6330	0.0000	0.0000	0.6747
MOLECULAR WEIGHT....	188.5288	0.0000	160.2012	0.0000	0.0000	188.7466
ACT.DENS LB /FT3	41.1235	0.0000	42.9231	0.0000	0.0000	41.1392
STD. API GRAVITY....	29.0728	0.0000	37.2206	0.0000	0.0000	29.0401
***** DRY BASIS *****						
RATE LB /HR	3724.1621	0.0000	9193.2695	0.0000	0.0000	3699.0215
MOLECULAR WEIGHT....	188.6310	0.0000	160.3039	0.0000	0.0000	188.8486
UOP K	11.2746	0.0000	11.4930	0.0000	0.0000	11.2738
FLASH POINT, DEG F	4.7693	0.0000	-40.7185	0.0000	0.0000	5.0140
CRIT. TEMP, F	709.1351	0.0000	649.0773	0.0000	0.0000	710.1290
CRIT. PRES, PSIA	290.5542	0.0000	328.0555	0.0000	0.0000	290.5071
***** VAPOR PHASE *****						
RATE LB /HR	30156.5742	30181.7188	20988.4336	7417.4619	36.2091	0.0000
ACT.RATE FT3/SEC	6.52	6.55	5.52	3.52	0.02	0.00
STD.RATE MM FT3/HR	0.89	0.89	0.87	0.81	0.00	0.00
CP, BTU /LB F	1.0329	1.0311	1.1607	2.1437	2.1437	0.0000
MOLECULAR WEIGHT....	12.8611	12.8709	9.1749	3.4704	3.4704	0.0000
ACT.DENS LB /FT3	1.2840	1.2796	1.0557	0.5857	0.5857	0.0000
COMPRESSIBILITY (Z).	1.0321	1.0320	1.0332	1.0430	1.0430	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
***** LIQUID PHASE *****						
RATE LB /HR	3724.1621	0.0000	9193.2695	0.0000	0.0000	3699.0215
ACT.RATE BBL/DAY	387.11	0.00	915.54	0.00	0.00	384.35
CP, BTU /LB F	0.6747	0.0000	0.6330	0.0000	0.0000	0.6747
MOLECULAR WEIGHT....	188.6310	0.0000	160.3039	0.0000	0.0000	188.8486
ACT.DENS LB /FT3	41.1234	0.0000	42.9225	0.0000	0.0000	41.1391
STD. API GRAVITY....	29.0739	0.0000	37.2223	0.0000	0.0000	29.0412
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	14	15	16	17
STREAM NAME..........	COLD SEP LIQ	REC COMP DIS	WASH H2O	SOUR H2O
STREAM PHASE.........	LIQUID	VAPOR	LIQUID	LIQUID
TEMPERATURE, DEG F	120.0000	152.8118	125.0000	120.0000
PRESSURE, PSIA	1095.0000	1265.0000	1110.0000	1095.0000
RATE LB MOLS/HR	197.5497	2140.6670	0.9704	0.5511
RATE LB /HR	22735.0977	7477.0352	17.4815	9.9296
ENTHALPY MM BTU /HR	0.4840	-2.4979	0.0017	0.0003
ENTHALPY BTU /LB	21.2877	-334.0800	95.7991	27.0891
MOLECULAR WEIGHT....	115.0854	3.4929	18.0150	18.0189
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	7477.0352	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	3.24	0.00	0.00
STD.RATE MM FT3/DAY	0.00	19.50	0.00	0.00
CP, BTU /LB F	0.0000	2.1395	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	3.4929	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.6413	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	1.0483	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	22735.0977	0.0000	17.4815	9.9296
ACT.RATE BBL/DAY	2107.75	0.00	1.21	0.69
STD. LV RATE BBL/HR	85.71	0.00	0.05	0.03
CP, BTU /LB F	0.5026	0.0000	0.9928	1.1685
MOLECULAR WEIGHT....	115.0854	0.0000	18.0150	18.0189
ACT.DENS LB /FT3	46.1073	0.0000	61.6306	61.4190
STD. API GRAVITY....	55.1029	0.0000	10.0635	10.5810
*** DRY BASIS ***				
RATE LB /HR	22727.9883	0.0000	0.0000	0.0557
MOLECULAR WEIGHT....	115.2796	0.0000	0.0000	18.7403
UOP K	11.9576	0.0000	0.0000	11.6004
FLASH POINT, DEG F	-91.1659	0.0000	0.0000	1.8953
CRIT. TEMP, F	523.3317	0.0000	0.0000	181.3166
CRIT. PRES, PSIA	404.9004	0.0000	0.0000	1402.2766
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	7417.5693	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	3.23	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.81	0.00	0.00
CP, BTU /LB F	0.0000	2.1530	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	3.4704	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.6371	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	1.0484	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	22727.9883	0.0000	0.0000	0.0557
ACT.RATE BBL/DAY	2107.26	0.00	0.00	0.01
CP, BTU /LB F	0.5023	0.0000	0.0000	1.5294
MOLECULAR WEIGHT....	115.2796	0.0000	0.0000	18.7403
ACT.DENS LB /FT3	46.1055	0.0000	0.0000	33.5432
STD. API GRAVITY....	55.1170	0.0000	0.0000	102.3169
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

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SIMULATION SCIENCES INC. PROCESS™ PAGE 19
PROJECT GP JET FUELS MAX JP4
PROBLEM C1F205 SOLUTION DEC 04 1987

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	25	26	27	28	30	31
STREAM NAME..........	HDC NAPHTHA	HDC JP-4	FRAC CVHD	HDC HC VAP	FRAC STR STM	JP-4 STR STM
STREAM PHASE........	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F	90.0000	262.8516	200.4344	90.0000	322.6034	312.2646
PRESSURE, PSIA	60.0000	55.3000	70.0000	60.0000	74.7998	55.3000
RATE LB MOLS/HR	39.9516	115.9792	425.3399	25.6295	33.3056	37.1912
RATE LB /HR	2853.6284	14037.0020	25687.5742	706.5347	599.9996	669.9996
ENTHALPY MM BTU /HR	0.0515	1.3050	6.6696	0.0994	0.7141	0.7974
ENTHALPY BTU /LB	18.0337	92.9706	259.6436	140.7007	1190.2109	1190.2109
MOLECULAR WEIGHT....	71.4272	121.0303	60.3930	27.5672	18.0150	18.0150
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	0.0000	25687.5742	706.5347	599.9996	669.[illegible]
ACT.RATE FT3/SEC	0.00	0.00	11.14	0.69	1.00	1.50
STD.RATE MM FT3/DAY	0.00	0.00	3.87	0.23	0.30	0.34
CP, BTU /LB F	0.0000	0.0000	0.5009	0.5283	0.5550	0.5307
MOLECULAR WEIGHT....	0.0000	0.0000	60.3930	27.5672	18.0150	18.0150
ACT.DENS LB /FT3	0.0000	0.0000	0.6403	0.2855	0.1674	0.1242
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9321	0.9820	0.9589	0.9685
*** LIQUID PHASE ***						
RATE LB /HR	2853.6284	14037.0020	0.0000	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY	311.46	1443.16	0.00	0.00	0.00	0.00
STD. LV RATE BBL/HR	12.64	52.79	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5444	0.5834	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	71.4272	121.0303	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	39.1644	41.5768	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	87.7928	54.6287	0.0000	0.0000	0.0000	0.0000
*** DRY BASIS ***						
RATE LB /HR	2352.8999	13990.8379	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	71.4813	123.3578	0.0000	0.0000	0.0000	0.0000
UOP K	12.7357	11.9769	0.0000	0.0000	0.0000	0.0000
FLASH POINT, DEG F	-66.2333	33.1789	0.0000	0.0000	0.0000	0.0000
CRIT. TEMP, F	383.9533	600.5282	0.0000	0.0000	0.0000	0.0000
CRIT. PRES, PSIA	503.1558	397.5057	0.0000	0.0000	0.0000	0.0000
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	0.0000	24472.5000	701.1978	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	9.27	0.68	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.14	0.01	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.5051	0.5290	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	68.3308	27.6789	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.7332	0.2868	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9217	0.9818	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0091	0.0108	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	2352.8999	13990.8379	0.0000	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY	311.41	1439.79	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5443	0.5959	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	71.4813	123.3578	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	39.1607	41.5373	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	87.7125	54.7753	0.0000	0.0000	0.0000	0.0000
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	32	33
STREAM NAME..........	SOUR H2O	REFLUX
STREAM PHASE.........	LIQUID	LIQUID
TEMPERATURE, DEG F	90.0000	90.0000
PRESSURE, PSIA	60.0000	60.0000
RATE LB MOLS/HR	66.8175	292.9413
RATE LB /HR	1203.7168	20923.9805
ENTHALPY MM BTU /HR	0.0698	0.3773
ENTHALPY BTU /LB	58.0156	18.0337
MOLECULAR WEIGHT....	18.0150	71.4272
*** VAPOR PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	1203.7168	20923.9805
ACT.RATE BBL/DAY	82.86	2293.73
STD. LV RATE BBL/HR	3.44	92.71
CP, BTU /LB F	0.9978	0.5444
MOLECULAR WEIGHT....	18.0150	71.4272
ACT.DENS LB /FT3	62.1002	39.1644
STD. API GRAVITY....	10.0635	87.7927
*** DRY BASIS ***		
RATE LB /HR	0.0000	20918.6406
MOLECULAR WEIGHT....	0.0000	71.4814
UOP K	0.0000	12.7357
FLASH POINT, DEG F	0.0000	-86.2333
CRIT. TEMP, F	0.0000	383.9538
CRIT. PRES, PSIA	0.0000	503.1559
*** VAPOR PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	0.0000	20918.6406
ACT.RATE BBL/DAY	0.00	2233.36
CP, BTU /LB F	0.0000	0.5443
MOLECULAR WEIGHT....	0.0000	71.4814
ACT.DENS LB /FT3	0.0000	39.1667
STD. API GRAVITY....	0.0000	87.4125
VISCOSITY, [illegible]	[illegible]	[illegible]

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	29	40	41	42	43	44
STREAM NAME.........	HDT NAP	NAP FEED	STAB OFFGAS	REFLUX	NAP PROD	REBOILER
STREAM PHASE........	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID	LIQUID
TEMPERATURE, DEG F	90.0000	89.8248	94.1650	94.1650	262.5658	226.5963
PRESSURE, PSIA	150.0000	144.0000	125.0000	125.0000	142.0000	141.5000
RATE LB MOLS/HR	44.8024	84.7563	7.1663	17.9156	77.5554	146.9895
RATE LB /HR	3209.6333	6063.4775	294.8259	951.0134	5767.9932	10470.0488
ENTHALPY MM BTU /HR	0.0086	0.0601	0.0616	0.0347	0.6457	0.9597
ENTHALPY BTU /LB	2.6839	9.9094	209.0047	36.4371	111.9418	91.6606
MOLECULAR WEIGHT....	71.6398	71.5397	41.1406	53.0829	74.3725	71.2299
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	0.0000	294.8259	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.08	0.00	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00	0.07	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.4772	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	41.1406	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.9812	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9818	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	3209.6333	6063.4775	0.0000	951.0134	5767.9932	10470.0488
ACT.RATE BBL/DAY	318.46	629.37	0.00	122.78	716.00	1275.42
STD. LV RATE BBL/HR	12.98	25.62	0.00	4.88	24.05	44.54
CP, BTU /LB F	0.5048	0.5233	0.0000	0.7108	0.6893	0.6920
MOLECULAR WEIGHT....	71.6398	71.5397	0.0000	53.0829	74.3725	71.2299
ACT.DENS LB /FT3	43.0820	41.1822	0.0000	33.1094	34.4351	35.0904
STD. API GRAVITY....	68.6129	77.6396	0.0000	122.6633	74.8996	79.0423
*** DRY BASIS ***						
RATE LB /HR	3208.8169	6061.9326	0.0000	950.6528	5767.8945	10468.9688
MOLECULAR WEIGHT....	71.6941	71.5939	0.0000	53.1222	74.3765	71.2516
UOP K	11.7920	12.2361	0.0000	13.8836	12.1437	12.2821
FLASH POINT, DEG F	-73.3390	-83.7349	0.0000	-118.4625	-57.6594	-77.2691
CRIT. TEMP, F	422.6771	404.6239	0.0000	263.5591	421.0440	401.0717
CRIT. PRES, PSIA	575.3867	541.3368	0.0000	603.4601	515.4408	529.5265
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	0.0000	294.0104	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.08	0.00	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.4775	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	41.2876	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.9856	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.8811	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0090	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	3208.8169	6061.9326	0.0000	950.6528	5767.8945	10468.9688
ACT.RATE BBL/DAY	318.40	629.26	0.00	122.76	716.00	1275.34
CP, BTU /LB F	0.5047	0.5232	0.0000	0.7107	0.6893	0.6920
MOLECULAR WEIGHT....	71.6941	71.5939	0.0000	53.1222	74.3765	71.2516
ACT.DENS LB /FT3	43.0787	41.1737	0.0000	33.1036	34.4349	35.0889
STD. API GRAVITY....	[illegible]	[illegible]	0.0000	122.7061	74.9007	79.0494
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

APPENDIX C

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 2
PROFITABLE JP-4 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

W16242/bf

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to produce JP-4 type aviation turbine fuel and chemical byproducts to maximize profit from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- Tar Oil byproduct stream (47620 #/hr, 3182 BPSD) is charged to the hydrotreater (Area 100).

- The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 500°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (3400 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.

- The hydrotreater produces 6 streams:

  - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

  - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

  - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to storage and gasoline blending.

  - JP-4 turbine fuel which is combined with JP-4 produced in the hydrocracker (area 200) and sent to storage.

  - 500°F+ unconverted bottoms product which is sent to the hydrocracker (area 200).

  - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

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- Approximately 950 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.

. The 500^0F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-4 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.

. The hydrocracker produces 4 streams in addition to JP-4

- High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H_2 and CH_4.

- Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

- Stabilized naphtha which is sent to storage and gasoline blending.

- A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.

. Hydrogen make-up for the Hydrotreater, the Hydrocracker and the Naphtha Hydrotreater (Area 600) is supplied from a PSA Hydrogen Unit (Area 300). High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft^3. This H_2, CO & CH_4 rich gas is recompressed into the methanation unit of the SNG plant.

. The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).

. The distillation removes the material boiling below 160^0F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.

. The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.

1.1 Overall Process Description - cont'd

- The naphtha hydrotreater produces 4 streams:

    - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

    - Naphtha which is stabilized to control vapor pressure. Approximately 74% of the naphtha is the sent to the Aromatics Recovery Unit (Area 700), the remainder is sent to gasoline blending.

    - A low pressure off gas which is sent to the Stretford unit in the SNG plant.

    - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

- A portion of the hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.

- Five streams are produced in the ARU plant.

    - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.

    - A small process water stream which is sent to the waste treatment plant in the SNG Plant.

    - Three product streams Benzene, Toluene & Xylene which are sent to storage.

- The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).

1.1 <u>Overall Process Description</u> - cont'd

- Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.

- The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.

- The acid tar produced is water washed and mixed with light oil and sent to fuel.

- The remaining cresol/xylenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).

- Streams produced in the phenol extraction unit are:

  - Phenol product sent to storage

  - Crude Cresylic Acid sent to distillation (Area 900) or storage.

  - Wash Water sent to Water Treatment in the SNG Plant.

  - Waste Water sent to the Phenosolvan unit in the SNG Plant.

  - Neutral Oil sent to storage and fuel for the SNG Plant boilers.

- The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.

- Streams produced in the crude cresylic acid distallation unit are:

  - o-Cresol product which is sent to storage.

  - m,p-Cresol product which is sent to storage.

  - Xylenol product which is sent to storage.

  - A heavy distillate which is combined with neutral oil in Area 800.

1.1 Overall Process Description - cont'd

- A crude phenol stream which is recycled to the Area 800.

- A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas, neutral oil and 160°F- distillate produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

936	BPSD of Crude Phenol
725	BPSD of Crude Naphtha
3182	BPSD of Tar Oil
4347	BPSD of #6 Fuel Oil
9.52	MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

3403	BPSD of JP-4 turbine fuel
324	BPSD of 160°F - Naphtha for gasoline blending
317	BPSD of Phenol
56	BPSD of o-Cresol
131	BPSD of m,p-Cresol
75	BPSD of Xylenols
312	BPSD of Neutral Oil for Fuel
202	BPSD of 160°F - Distillate for Fuel
161	BPSD of Gasoline Blending Stock
233	BPSD of Benzene
83	BPSD of Toluene
11	BPSD of Xylene
6.27	MMSCFD equivalent SNG product credit due to HDT, HDC & PSA purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	4347 BPSD
SNG Equivalent of Syn Gas & Purge Gas	3.25 MM SCFD
Power	6180 kW
Cooling Water	6140 GPM (30°F rise)
Process Water	31.0 GPM

In addition the process utilizes steam as summarized below which was debited against boiler requirements.

HP Steam Import	54,700 #/H
MP Steam Import	8,900 #/H
LP Steam Export	6,900 #/H
Condensate Return	63,600 #/H

Figure 1 Case 2:Profitable JP-4

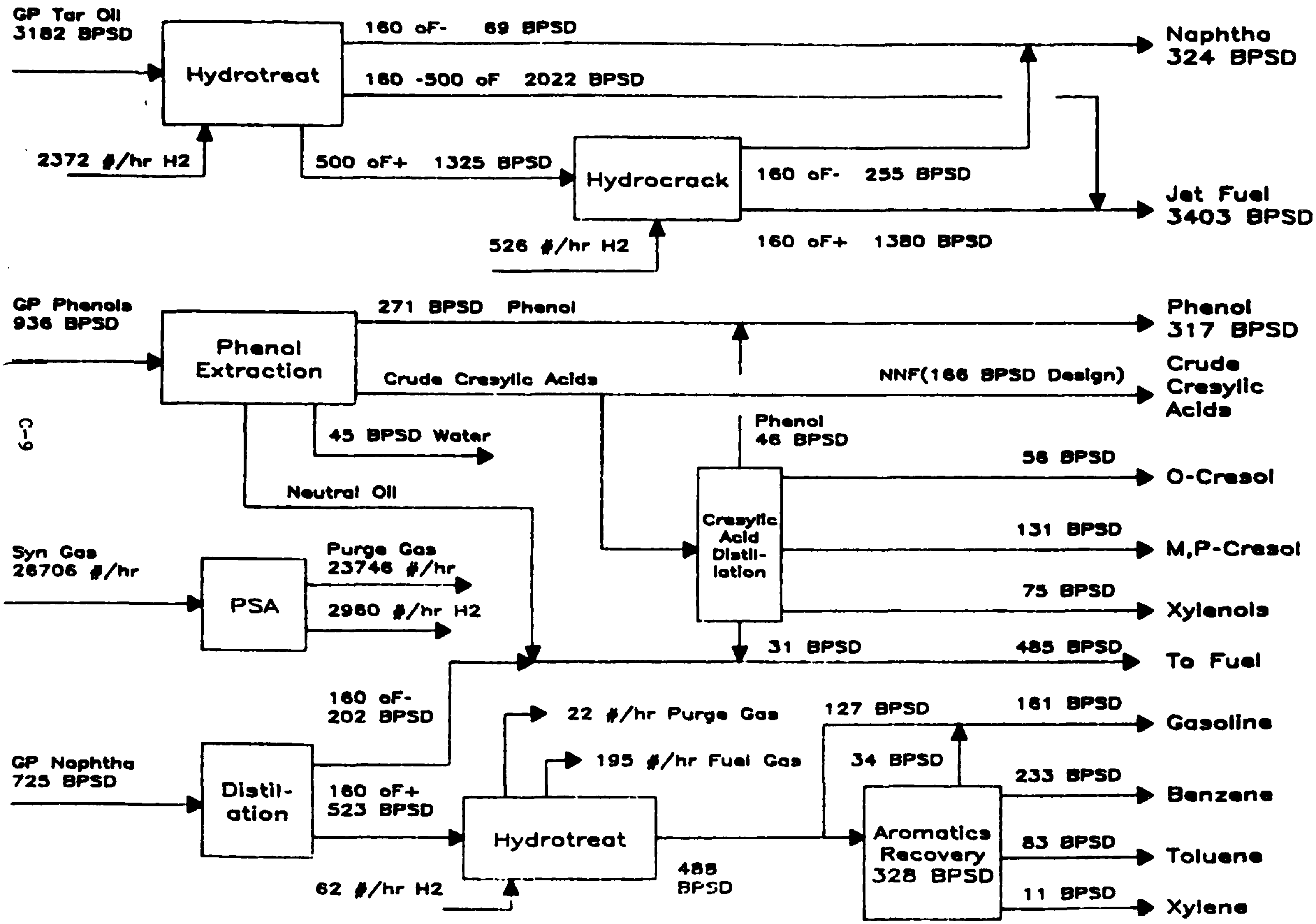


Table 1.1 Great Plains Case 2: Profitable JP4 Production

Tar Oil Feed=====>	47620	#/hr	3182	BPSD
Phenol Feed======>	14490	#/hr	936	BPSD
Crd Naphtha Feed=>	8738	#/hr	725	BPSD
Naphtha Product==>	3370	#/hr	324	BPSD
JP4 Product======>	40117	#/hr	3403	BPSD
Phenol Product===>	4925	#/hr	317	BPSD
o-Cresol Prod====>	845	#/hr	56	BPSD
m,p-Cresol Prod==>	1974	#/hr	131	BPSD
Xylenols Prod====>	1070	#/hr	75	BPSD
Gasoline Stock===>	1957	#/hr	161	BPSD
Benzene Prod=====>	3003	#/hr	233	BPSD
Toluene Prod=====>	1054	#/hr	83	BPSD
Xylene Prod======>	139	#/hr	11	BPSD
SNG Product Loss=>	5586	#/hr	3.3	MMSCFD
Fuel Oil Makeup==>	60183	#/hr	4347	BPSD

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feeds					
H2	4.98		2373	1176.9	
Tar Oil	100.00	1.0268	47620		3182
Total	104.98		49993		
Products					
Purge Gas	0.11		54	16.8	
Fuel Gas	1.99		950	46.8	
Naphtha	1.61	0.6825	765		77
JP-4	50.86	0.8218	24221		2022
500 oF+	39.95	0.9850	19026		1325
H2O in SW	8.96		4267	237.1	
H2S in SW	0.43		205	6.0	
NH3 in SW	1.06		505	29.7	
Total	104.98		49993		3424

Fixed Bed Hydrocracker

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feeds					
H2	2.76		526	260.9	
500oF+	100.00	0.9850	19026		1325
Total	102.76		19552		
Products					
Purge Gas	0.94		179	60.7	
Fuel Gas	4.31		820	45.3	
Naphtha	13.96	0.6675	2656		273
JP-4	83.55	0.7900	15896		1381
H2S in SW	0.003		0.6	0.02	
NH3 in SW	0.003		0.6	0.03	
Total	102.77		19552		1654

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
HDT Nap	22.37	0.6825	765		77
HCR Nap	77.63	0.6675	2656		273
Stab Nap	98.50	0.7140	3370		324
Fuel Gas	1.50		51	1.2	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1707.6	503.1	39.9	438.2	8.5	5.1	2702.6
Prod. H2	1468.5	0.1	0.0	0.0	0.0	0.0	1468.7
Purge Gas	239.1	503.0	39.9	438.2	8.5	5.1	1233.9
#/hr							
Feed Gas	3443	14091	1760	7030	255	164	26743
Prod. H2	2961	4	0	0	0	0	2964
Purge Gas	482	14087	1760	7030	255	164	23779

Crude Naphtha Distillation

	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

Aromatics Recovery

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	4552	361
Products				
Raffinate	7.82	0.7175	356	34.0
Benzene	65.97	0.8844	3003	233.0
Toluene	23.15	0.8718	1054	83.0
Xylene	3.05	0.8729	139	10.9
Total Products	100		4552	361

Phenol Extraction

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Cr. Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m.p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	950	46.8	17.1
HCR FG Produced	820	45.3	14.8
Stabilizer FG	51	1.2	0.9
Naphtha Hdtr FG	195	10.8	3.5
Total Fuel Gas	1821	93.3	32.8

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	482	239.1	324	29.4
CO	14087	503.0	321	61.2
CO2	1760	39.9	0	0.0
C1	7030	438.2	1010	167.7
C2	255	8.5	1769	5.7
N2+Ar	164	5.1	0	0.0
Total	23779	1233.9	565	264.0

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BFSD
Tar Oil	-47620	17000	-809.5			-3182
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
Fuel Gas	1821	18000	32.8	0.8	927	
160 oF- distillate	2164	17400	37.7			202
Neutral Oil	4936	15000	74.0			312
Import Steam	-56700	1000	-56.7			
Fuel Oil to Boiler	59600	18000	1072.8			4305
Total	-59027		0.0	0.8		-24
Fuel Oil to Process Heaters	583	18000	10.5			42

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	9.52	64862
SNG Credit for PSA Purge gas	5.96	29613
SNG Credit for Hdtrs purge gas	0.31	2023
Total SNG Production Loss	3.25	

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

For a description of the Hydrotreater process see Case 3 Section 2.1.

2.2 Hydrocracker (Area 200)

For a description of the Hydrocracker process see Case 1 Section 2.2.

2.3 PSA Hydrogen Unit & Recompression (Area 300)

2.3.1 Hydrogen for both hydrotreaters and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 ^{o}F
Composition	mol%

H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

The PSA unit selectively absorbs all components expect H2 and produces a 99.99% vol. purity stream at about 345 psig and 80^{o}F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100^{o}F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

2.3 PSA Hydrogen Unit & Recompression (Area 300) - cont'd

2.3.1 Cont'd

The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-20301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

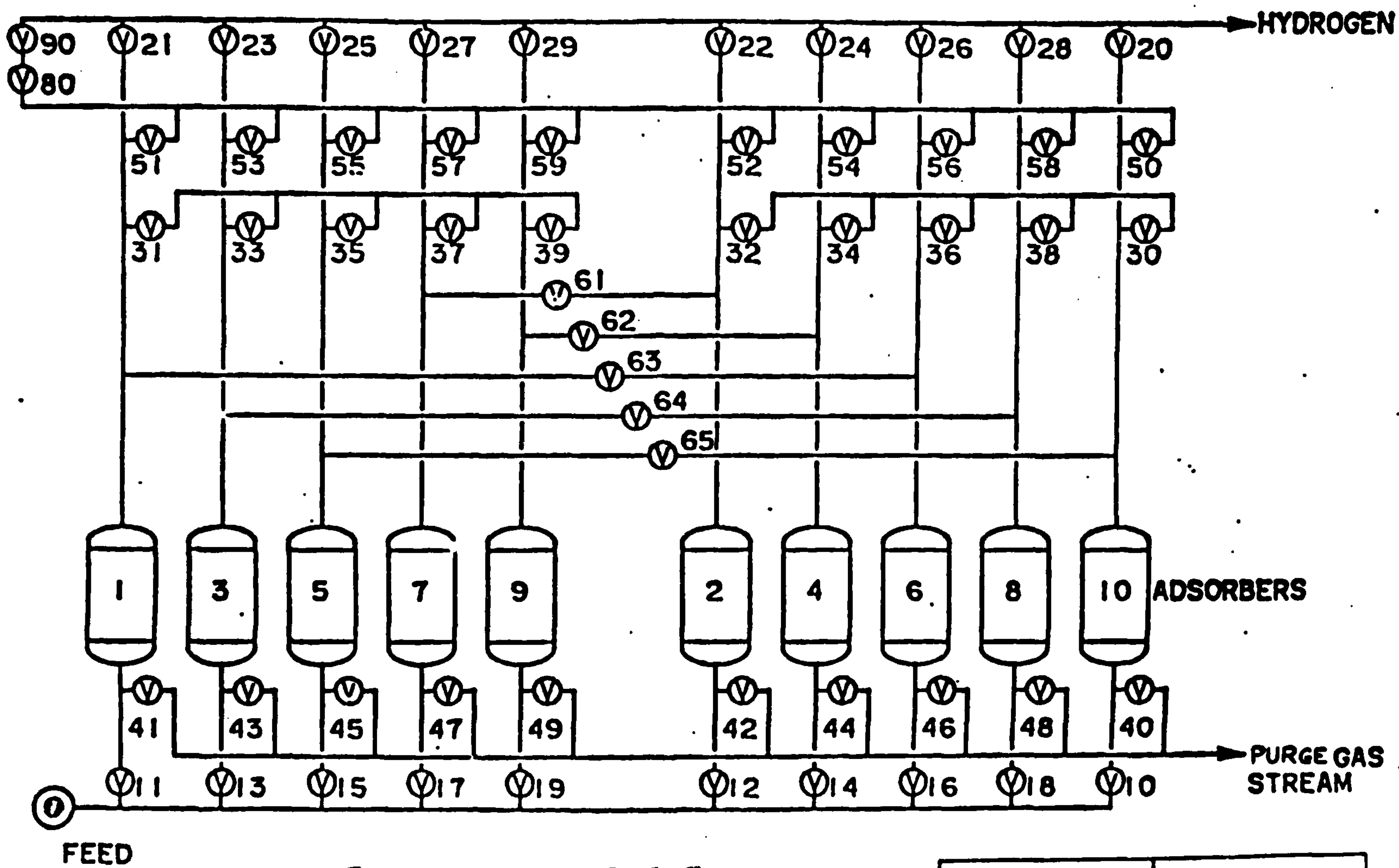
2.4 Phenol Stream (Area 800 and 900)

For description of the Phenol Extraction (Area 800) and Cresylic Acid Distillation (Area 900) Units see Case 7 Section 2.1.

2.5 Naphtha Stream (Areas 600 and 700)

For a description of the Naphtha Distillation and Hydrotreating Unit (Area 600) and the Aromatics Recovery Unit (Area 700) see Case 7 Section 2.2.

SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 10

REV	ISSUE DATE	DESCRIPTION	PROC ENGR	PROJ ENGR	APPR	APPR
△						
△	1/3/...	FOR SUBTASK 1.2	ML	E.S.		

LUMMUS — THE LUMMUS COMPANY

TITLE PSA HYDROGEN UNIT AREA 300

CLIENT AMOCO/DOE

PROJ NO 5571

CASE 2

DWG SKB - 5571-20301

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 2 - PROFITABLE JP-4

AREA 100 - HYDROTREATER

TAG. NO.	DESCRIPTION
See Case 3 Area 100	

AREA 200 - HYDROCRACKER

See Case 1 Area 200

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

AREA 400 - STORAGE AREA

FB-401	Jet Fuel Storage Tank
FB-402	Naphtha Storage Tank
FB-403	Fuel Oil Storage Tank
FB-404	Blending Stock
FB-405	Benzene Storage
FB-406	Toluene Storage
FB-407	Xylene Storage
FB-409	Gasoline Storage
FB-410	Neutral Oil Storage
FB-411	Phenol Product Storage
FB-412	Crude Cresylic Acid Storage
FB-413	O-Cresol Storage
FB-414	M, P Cresol Stoage
FB-415	Xylenol Storage

3.0 CAPITAL COSTS

3.1 Equipment List - cont'd

CASE 2 - PROFITABLE JP-4

TAG NO.	DESCRIPTION
AREA 400	STORAGE AREA
GA-401A/S	Tar/Tar Oil Feed Pump
GA-402A/S	Crude Phenol Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump
GA-405A/S	Crude Naphtha Transfer Pump
GA-406A/S	Blending Stock Pump
GA-407A/S	Benzene Transfer Pump
GA-408A/S	Toluene Transfer Pump
GA-409A/S	Xylene Transfer Pump
GA-411A/S	Gasoline Transfer Pump
GA-412A/S	Neutral Oil Transfer Pump
GA-413A/S	Crude Cresylic Acid Transfer Pump
GA-414A/S	O-Cresol Transfer Pump
GA-415A/S	M, P. Cresol Transfer Pump
GA-416A/S	Xylenol Transfer Pump
PA-401	Gasoline Blending Package

AREA 500 - CATALYST HANDLING

See Case 3 Area 500

AREA 600 - NAPHTHA DISTILL. AND HDT

See Case 7 Area 600

AREA 700 - AROMATICS RECOVERY

See Case 7 Area 700

AREA 800 - PHENOL EXTRACTION

See Case 7 Area 800

AREA 900 - CRESYLIC ACID DISTILLATION

See Case 7 Area 900

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate for this case is a factored type estimate using the T.I.C. values developed for the various cases referenced in this project.

The total investment costs are scaled to the capacity requirement of this case using a o.6 exponent.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs for Areas 100 thru 700 and 30% for Areas 800 & 900.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of $)

	Case 2
Area 100 Hydrotreater	$20,702
Area 200 Hydrocracker	10,430
Area 300 PSA & Recompression	8,100
Area 400 OSBL	8,443
Area 500 Catalyst Handling	1,285
Area 600 Naph. Dist & HDT	4,615
Area 700 ARU	7,887
Area 800 Phenol Ext.	12,276
Area 900 Cresylic Acid Dist.	4,832
Subtotal	$78,570
Area 700 ARU Solvent Inventory	80
Total	$78,650

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

This unit has the same capacity as the 100 Area of Case 3. Therefore, T.I.C. = $20,702.

Area 200

This unit has a 107% capacity of the 200 Area of Case 1. Therefore, T.I.C. = $(1.07)^{0.6}$ (10,012) = $10,430

3.2.3 Estimate Breakdown - Cont'd

Area 300

This unit has a 96% capacity of the 300 Area of Case 3. Therefore T.I.C. = $(0.96)^{0.6}$ (8300) = $8100.

Area 400

Tar Oil Stream Storage 80% Case 1
TIC = $(0.8)^{.6}$ (5 m) = $4300

Phenol Stream Storage = 100% Case 7
TIC = $3,016

Naphtha Stream Storage 75% Case 7
TIC = $(0.75)^{.6}$ (3058) = $2,890
Subtotal 10,206

Less Duplicate Pipe & Rack -1763
Total = $8,443

Area 500

The capacity of this unit is identical to the 500 Area of Case 3. Therefore T.I.C. = $1,285

Area 600

This unit has a capacity of 100% of the 600 Area of Case 7. Therefore T.I.C. = $4,615

Area 700

This unit has a capacity of 75% of the 700 Area of Case 7. Therefore T.I.C. = $(0.75)^{0.6}$ (9,373) = $7,887

Area 800

This unit has a capacity of 100% of the 800 Area of Case 7. Therefore T.I.C. = $ 12,276

Area 900

This unit has a capacity of 100% of the 900 Area of Case 7. Therefore = T.I.C. = $ 4,832

W16242/bf

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require men/shift to operate the plant broken down as follows:

Foreman	2
Control Room	2
HDT Operator	2
HCR Operator	2
PSA & relief man	1
Naphtha Distil. & HDT	2
ARU	2
Phenol Ext.	1
Cresylic Acid Dist.	1
	15 Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 7 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	15 positions x 4 people/position -	60
Supervisor & Admin.		6
QC Technician		2
Maintenance		7
Other (Stores or Janitorial)		1
Total		76

4.2 Utilities

The following utilities have been estimated:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	4347 BPSD	$16/Bbl [(a)]	69,552
SNG equivalent of Syn Gas & Purge Gas	3.25 MMSCFD	$3.80/MM Btu [(b)]	12,105
Cooling Water	6140 GPM	$0.155/MGal [(c)]	1,370
Power	6180 kW	$0.04/kWH [(c)]	5,933
Process Water	31 GPM	$0.45/MGal [(c)]	20

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

4.2 Utilities - cont'd

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

Catalyst & Chem.	Use	Cost	$/SD
Nap. HDT Cal	0.021 #/Bbl	$3.00/#	33
HDT Cat.	0.30 #/Bbl	$3.00/#	2864
HCR Cat.	0.013 #/Bbl	$6.00/#	104
Inhibitors	50 PPM	$10/Gal	92
ARU Solvent	18 #/D	$2.10/#	38
H_2SO_4	7100 #/D	$0.04/#	285
			$3416

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units (excluding ARU solvent inventory). On this basis the maintenance supplies would be

$$0.00005 \times 78,570,000 = \$\ 3929/SD$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-4 and by-product chemicals are proposed to be located to the east of the Phenosolvan Unit and Water Treatment Area of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 340′ x 575′ will be surrounded by an access road and will be divided by three central east-west roads. Areas 100 & 500 will be located to the north and Areas 200 & 300 South of Area 100, Areas 800 & 900 next and then Areas 600 & 700.

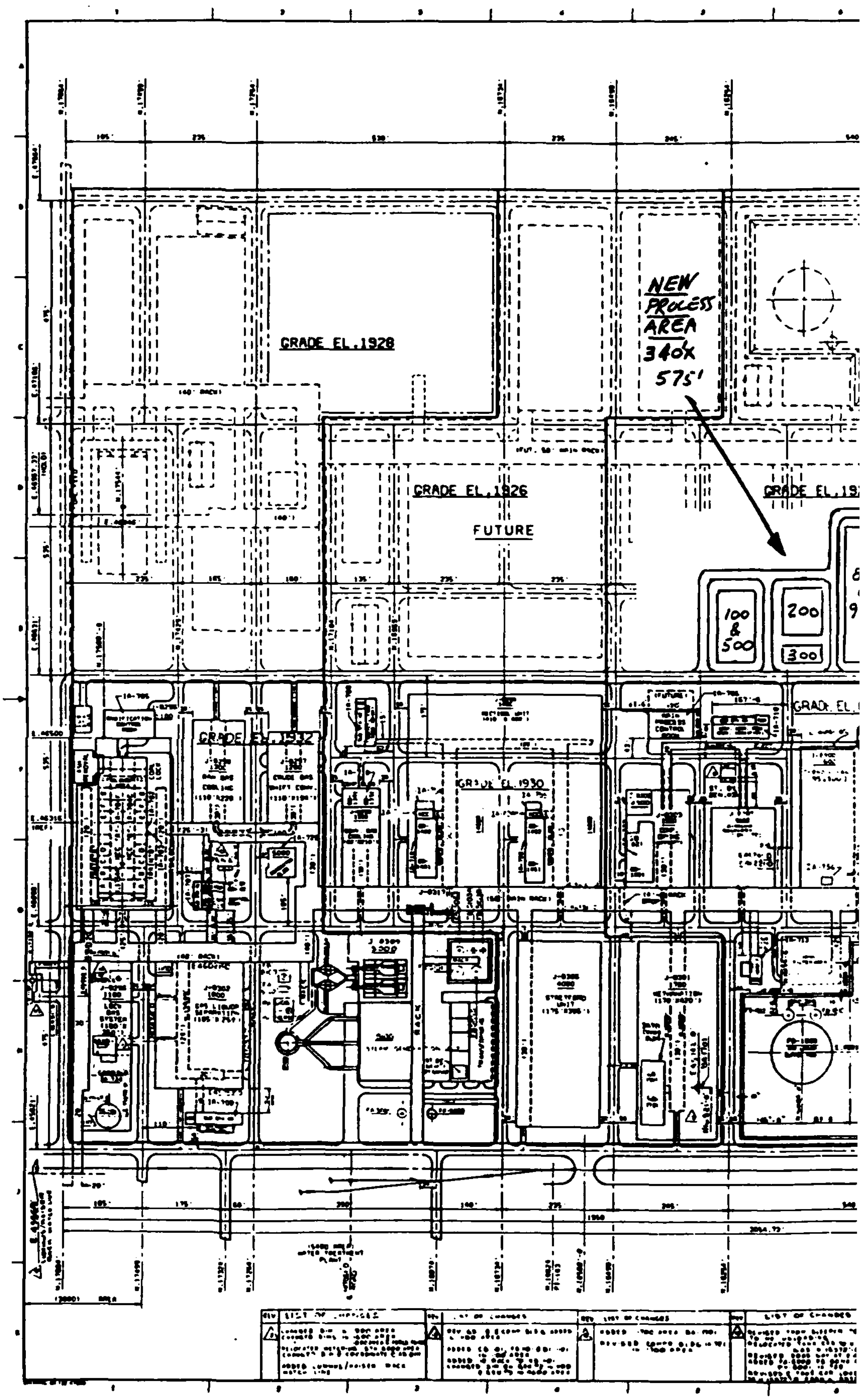
A diked storage tank area approx. 375′ x 375′ will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

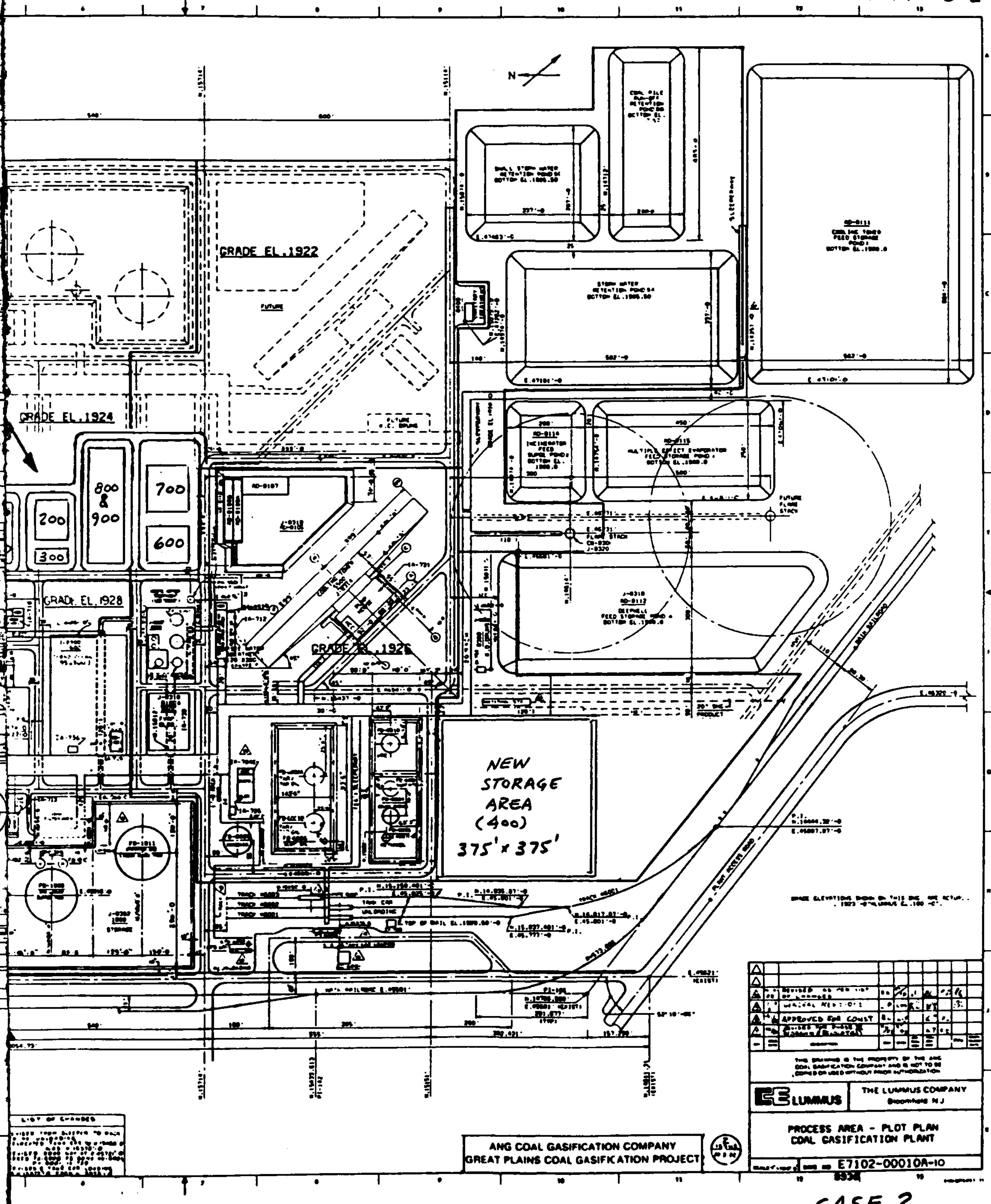
Approximately 2500 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

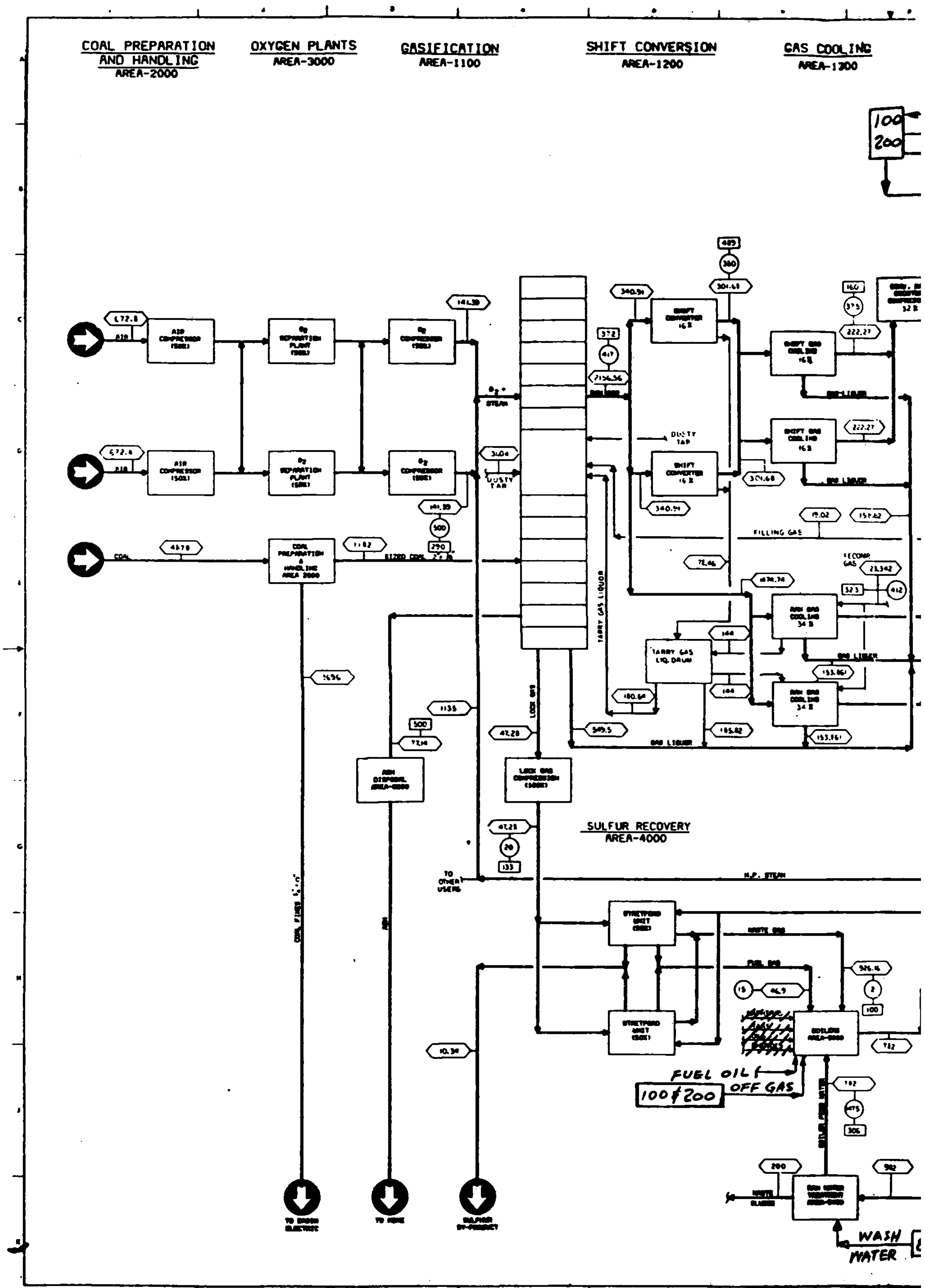
A summary of the lines has not been prepared for this case but will be similar to a combination of Cases 3 and 7 with the utility lines of like services being combined.

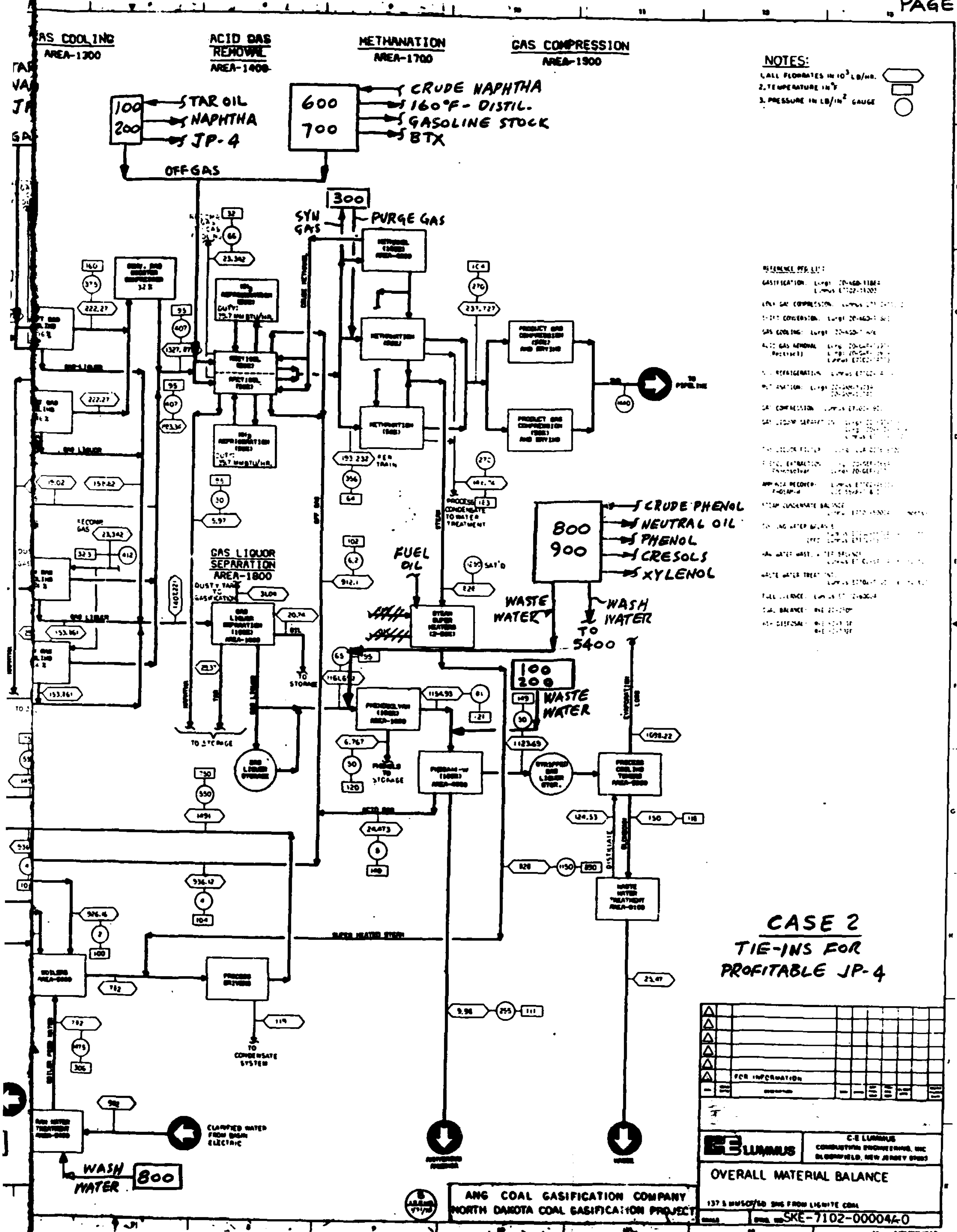


GRADE EL.1928
NEW PROCESS AREA 340'x 575'
GRADE EL.1926
FUTURE
100 & 500
200
300
GRADE EL.1932
GRADE EL.1930


GRADE EL.1922
GRADE EL.1924
GRADE EL.1928
GRADE EL.1926
800 & 900
700
200
600
300
NEW STORAGE AREA (400) 375'x375'
APPROVED FOR CONST
THE LUMMUS COMPANY
Bloomfield N J
LUMMUS
PROCESS AREA - PLOT PLAN
COAL GASIFICATION PLANT
E7102-0001DA-10
ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT

CASE 2


COAL PREPARATION AND HANDLING
AREA-2000
OXYGEN PLANTS
AREA-3000
GASIFICATION
AREA-1100
SHIFT CONVERSION
AREA-1200
GAS COOLING
AREA-1300
SULFUR RECOVERY
AREA-4000
100
200
FUEL OIL
OFF GAS
100 & 200
WASH
WATER
AIR
COAL
DUSTY TAR
TARRY GAS LIQUOR
FILLING GAS
GAS LIQUOR
H.P. STEAM
TO OTHER USERS
TO UNION ELECTRIC
TO MINE
SULFUR BY-PRODUCT



GAS COOLING
AREA-1300
ACID GAS REMOVAL
AREA-1400
METHANATION
AREA-1700
GAS COMPRESSION
AREA-1900
NOTES:
1. ALL FLOWRATES IN 10³ LB/HR.
2. TEMPERATURE IN °F
3. PRESSURE IN LB/IN² GAUGE
100
200
STAR OIL
NAPHTHA
JP-4
600
700
CRUDE NAPHTHA
160°F - DISTIL.
GASOLINE STOCK
BTX
OFF GAS
300
SYN GAS
PURGE GAS
GAS LIQUOR SEPARATION
AREA-1800
FUEL OIL
800
900
CRUDE PHENOL
NEUTRAL OIL
PHENOL
CRESOLS
XYLENOL
WASTE WATER
WASH WATER
TO 5400
100
200
WASTE WATER
TO STORAGE
TO PIPELINE
SUPER HEATED STEAM
TO CONDENSATE SYSTEM
CLARIFIED WATER FROM BASIN ELECTRIC
WASH WATER
800
CASE 2
TIE-INS FOR PROFITABLE JP-4
FOR INFORMATION
C-E LUMMUS
OVERALL MATERIAL BALANCE
ANG COAL GASIFICATION COMPANY
NORTH DAKOTA COAL GASIFICATION PROJECT
SKE-7102-000040

APPENDIX D

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 3
MAXIMUM JP-8 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

W16214/bf

TABLE OF CONTENTS

1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to maximize the production of type JP-8 aviation turbine fuel from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- Tar Oil byproduct stream (47620 #/hr, 3182 BPSD) is charged to the hydrotreater (Area 100).

- The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 550°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (4075 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.

- The hydrotreater produces 6 streams:

  - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

  - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

  - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to the main boiler in the SNG plant.

  - JP-8 turbine fuel which is combined with JP-8 produced in the hydrocracker (area 200) and sent to storage.

  - 550°F+ unconverted bottoms product which is sent to the hydrocracker (area 200).

  - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

- Approximately 950 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.

. The 550°F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-8 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.

. The hydrocracker produces 4 streams in addition to JP-8

- High pressure purge as (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H_2 and CH_4.

- Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

- Stabilized naphtha which is sent to the main boiler in the SNG plant.

- A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.

. Hydrogen make-up for both the Hydrotreater and the Hydrocracker is supplied from a PSA Hydrogen Unit. High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft^3. This H_2, CO & CH_4 rich gas is recompressed into the methanation unit of the SNG plant.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. Detailed material balances for each unit can be found in appendixes A&B. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas and naphtha produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

3182 BPSD of Tar Oil
1978 BPSD of #6 Fuel Oil
11.07 MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

2490 BPSD of JP-8 turbine fuel
7.37 MMSCFD equivalent SNG product credit due to HDT, HDC & PSA purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	1978 BPSD
SNG Equivalent of Syn Gas & Purge Gas	3.7 MM SCFD
Power	5450 kW
Cooling Water	2040 GPM (30^{o}F rise)
Process Water	20.5 GPM

In addition the process exports 6945 #/hr of 100 psig saturated steam which was credited against boiler requirements.

W16214/bf

Figure 1:Case 3-Maximum JP-8 Production

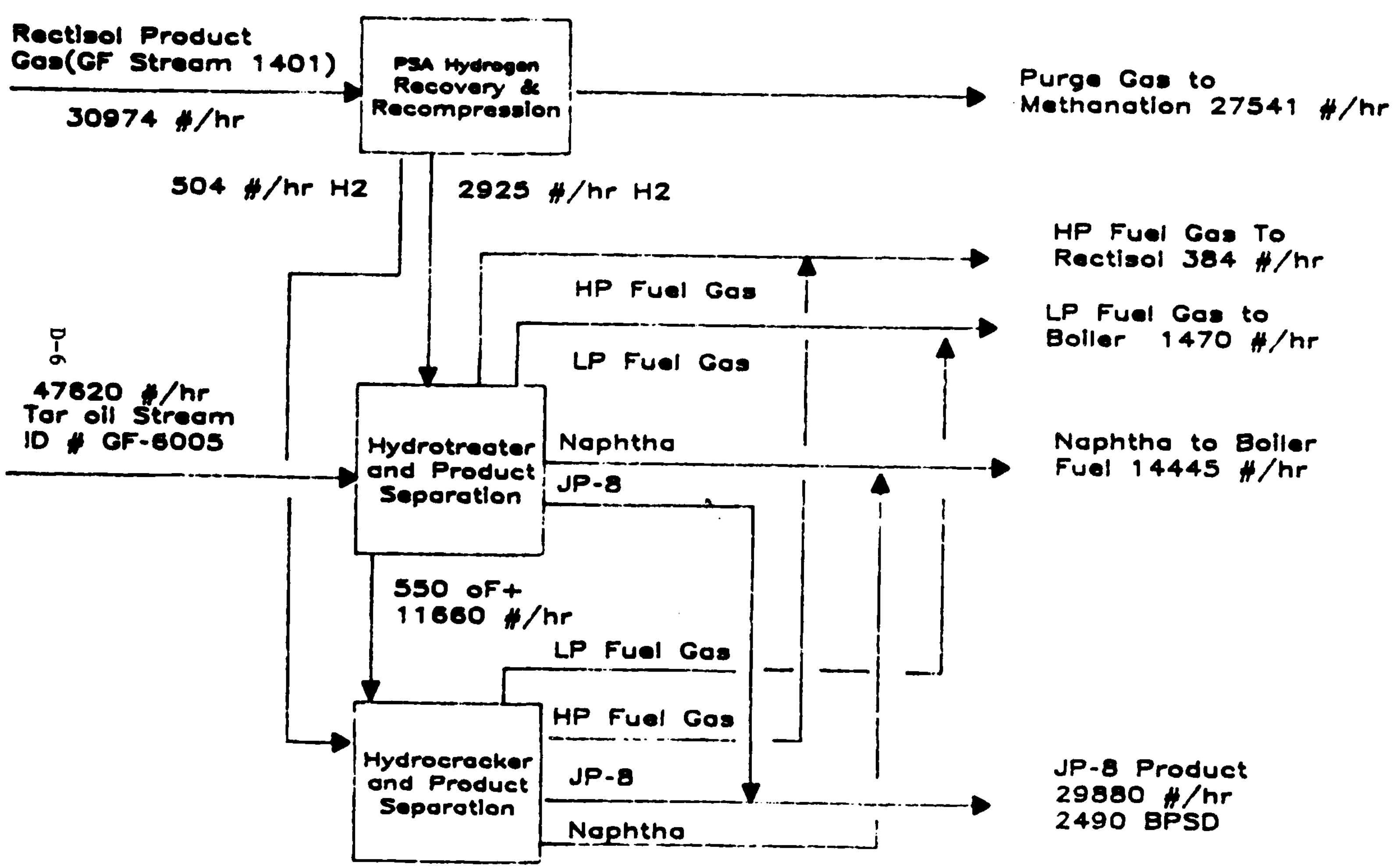


Table 1.3: Great Plains Case 3: Maximum JP8 Production

Tar Oil Feed=====>	47620	#/hr	3182	BFSD
JP-8 Product=====>	29880	#/hr	2490	BFSD
SNG Product Loss=>	6275	#/hr	3.7	MMSCFD
Fuel Oil Makeup==>	27386	#/hr	1978	BFSD

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BFSD
Feeds					
H2	6.14		2925	1450.9	
Oil	100.00	1.0268	47620		3182
Total	106.14		50545		
Products					
Purge Gas	0.10		48	14.9	
Fuel Gas	1.87		891	43.9	
Naphtha	19.17	0.7606	9127		823
JP-8	50.29	0.8220	23950		1975
550 oF+	24.49	0.9697	11660		825
H2O in SW	8.73		4159	231.1	
H2S in SW	0.43		205	6.0	
NH3 in SW	1.06		505	29.7	
Total	106.14		50545		3623

Fixed Bed Hydrocracker

Comp.	Wt %	Grav	#/hr	#Mole/hr	BFSD
Feeds					
H2	4.32		504	250.0	
550oF+	100.00	0.9697	11660		825
Total	104.32		12164		
Products					
Purge Gas	2.88		336	113.9	
Fuel Gas	2.12		247	13.7	
Naphtha	48.46	0.7148	5650		542
JP-8	50.86	0.7900	5930		515
H2S in SW	0.003		0.4	0.01	
NH3 in SW	0.003		0.4	0.02	
Total	104.32		12164		1057

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BFSD
HDT Nap	61.76	0.7606	9127		823
HCR Nap	38.24	0.7148	5650		542
Stab Nap	97.75	0.7346	14445		1349
Fuel Gas	2.25		332	7.8	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	6.63	5.55	901.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1977.8	582.7	46.3	507.5	9.9	6.0	3130.2
Prod. H2	1700.9	0.2	0.0	0.0	0.0	0.0	1701.1
Purge Gas	276.9	582.6	46.3	507.5	9.9	6.0	1429.1
#/hr							
Feed Gas	3987	16320	2038	8143	296	190	30974
Prod. H2	3429	4	0	0	0	0	3433
Purge Gas	558	16316	2038	8143	296	190	27541

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	891	43.9	16.0
HCR FG Produced	247	13.7	4.4
Stab FG Produced	332	7.8	6.0
Total Fuel Gas	1470	65.4	26.5

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	558	276.9	324	34.0
CO	16316	582.6	321	70.9
CO2	2038	46.3	0	0.0
C1	8143	507.5	1010	194.3
C2	296	9.9	1769	6.6
N2+Ar	190	6.0	0	0.0
Total	27541	1429.1	565	305.8

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Tar Oil	-47620	17000	-809.5			-3182
Fuel Gas	1470	18000	26.5	0.6	1068	
Naphtha	14445	20040	289.5			1349
Export Steam	6945	1000	6.9			
Fuel Oil to Boiler	27036	18000	486.7			1753
Total	2277		0.0	0.6		12[illegible]
Fuel Oil to Process Heaters	350	18000	6.3			25

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	11.02	75124
SNG Credit for PSA Purge gas	6.90	34[illegible]98
SNG Credit for Hdr. purge gas	0.47	3091
Total SNG Production Loss	3.65	

Water Balance Hydrotreater

Component	H2O (#/hr)	H2S	NH3	NH4HS	NH4OH	Total
Reaction Gases	4159	205	505			4869
Reaction Solution	3733			308	829	4869
Stripping Steam	1225					1225
Softened Water	9281					9281
HDT Sour Water	14239			308	829	15[illegible]75

Water Balance Hydrocracker

Component	H2O (#/hr)	H2S	NH3	NH4HS	NH4OH	Total
Reaction Gases		0.4	0.4			0.7
Reaction Solution				0.5	0.4	0.9
Stripping Steam	225.1					225.1
HCR Sour Water	225.1			0.5	0.4	226.0

Total Sour Water

	H2O (#/hr)	H2S	NH3	NH4HS	NH4OH	Total
Total Sour Water	14464			308	829	15601

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

Operating conditions for the hydrotreater were provided to Lummus by Amoco and these conditions are presented in Table 2.1. The basic processing step selected was the expanded bed hydrotreater (LC Fining) system. Due to the extremely high exothermic heat of reaction it was necessary to use 3 reactors in with interstage cooling. Referring to drawing D5571-30101 and the material balance printouts (Appendix A) the flow is as follows:

- Feed Tar Oil is charged into the hydrotreater from day tank FA-101 and through charge pump GA-101 and preheater exchanger EA-101.
- The preheated charge oil is combined with feed hydrogen gas from heater BA-101. Preheat of the oil is limited to 550°F to prevent cracking. The preheated mixture is then charged to the first reactor DC-101A.
- The expanded bed reactor DC-101A approaches isothermal conditions in which the heat of reaction is used to heat the feed up to 700°F.
- The effluent from DC101A is cooled in exchanger EA-101 and combined with recirculating hydrogen from recycle hydrogen gas compressor GB-102. The combined mixture is charged into the second reactor where the heat of reaction increases the temperatures to 700°F.
- The effluent from DC101B is cooled in exchanger EA-104 and combined with recirculating hydrogen gas from recirculating compressor GB-102. This mixture is charged into the third reactor DC-101C.
- The effluent from DC101C goes to the high temperature/high pressure separator FA-103. Hot liquid from FA-103 flows to the hydrotreator fractionation DA-101. The vapors from FA-103 flows through exchangers EA-102 and EA-105 and then through air cooler EC-101. Process water is injected prior to EC-101 to convert the H2S and NH3 in the gas to an aqueous NH4OH/NH4HS solution.
- Exchangers EA-104 and EA-105 are part of a circulating hot oil belt which allows for the generation of steam from waste heat in the high pressure loop without having the problem of a hydrocarbon leak from the high pressure system into the steam system.

W16214/bf

2.1 Hydrotreater - Cont'd

- The cooled gas then passes into the High Pressure/Low Temperature Separator FA-104 where hydrogen rich gas is taken as an overhead product. A purge stream of this high pressure gas is taken (to purge H2 and light gases from the loop) and sent to the Rectisol Unit 1400 in the SNG plant to recover the hydrogen in the purge gas. The remaining gas is recirculated to reactors DC-101B and DC-101C.

- The water phase from separator FA-104 goes to the PHOSAM Unit in the SNG plant to recover the H2S and NH3.

- The hydrocarbon phase from separator FA-104 is preheated in exchanger EA-105 and is combined with the hot liquid from FA-103 and charged to the HDT Fractionator DA-101. Fractionator DA-101 produces 550°F+ product (which is sent to hydrocracking, area 200), JP-8 (which is sent to storage), and unstabilized naphtha (which is sent to the naphtha stabilizer in the hydrocracking area 200).

- Catalyst is replaced every third day in each reactor so that one reactor is receiving and withdrawing catalyst each day. Catalyst is added and replaced by the catalyst handling system.

- Waste heat is converted to 100 psig saturated steam in exchangers EA-107 and EA-108. This steam is used for stripping in DA-101 and in the hydrocracking area 200 for stripping steam. There is an excess of about 6945 #/hr which is exported to the SNG steam system.

Table 2.1 Hydrotreater Conditions

Case 3 Maximum JP-8 Operation

Reactor Type	Expanded Bed
Number of Reactors	3
Reactor Temperature	700°F
Temp. rise/stage	225°F max.
Ratio of H2 in Feed to Chemical H2	2.0 min.
Catalyst Replacement	0.30 #/Bbl

W16214/bf

2.2 <u>Hydrocracker</u> (Area 200)

Operating conditions for the hydrocracker were provided to Lummus by AMOCO and these conditions are presented in Table 2.2. The basic processing step selected was a 5 bed fixed bed reactor system with recycle of unconverted 550°F+ material. Beds 1 and 2 use a catalyst that is most active for sulfur, nitrogen and oxygen removal while beds 3,4,5 use a catalyst that is most active for hydrocracking. Referring to drawing D5571-30201 and the material balance printouts (Appendix B) the flow is as follows:

- Hydrotreated 550°F+ material from the hydrotreater (Area 100) enters the system from day tank FA-201 through feed pump GA-201 and is preheated in exchanger EA-201. The preheated feed is combined with unconverted bottoms from fractionator DA-201 (approximately 35% of the feed is recycled). The combined oils are then mixed with hot hydrogen coming from heater BA-201 and charged to the reactor.

  The combined feed to the first bed in the reactor is 662°F.

- In the first bed the temperature rises to about 675°F. Quench hydrogen is added to cool the effluent from the first bed to about 648°F. In each of the remaining beds quench hydrogen is added to cool the beds. The inlet and outlet temperatures from each bed are as follows:

	Inlet	Outlet
Bed 1	662	675
Bed 2	648	676
Bed 3	652	677
Bed 4	656	680
Bed 5	661	684

- The reactor effluent passes into the high temperature/high pressure separator FA-202. Vapors from FA-202 are cooled in EA-201, EA-202 and then air condenser EC-201. Water is injected into the condenser EC-201 to dissolve H2S and NH3 into a NH4OH/NH4HS solution. This solution is sent to the PHOSAM unit in the SNG plant.

- The cooled vapors pass into separator FA-203 and the overhead hydrogen rich gas is divided with the major portion being used as recycle gas to the reactors via compressor GB-202 and heater BA-201. A small portion of the gas is purged from the system as high pressure purge gas which goes to the Rectisol unit in the SNG plant.

2.2 Hydrocracker (Area 200)

- The hydrocarbon phase from separator FA-203 is heated in exchanger EA-202, combined with the hot oil from separator FA-202 and charged to fractionator DA-201.

- Fractionator DA-201 produces unstabilized naphtha (which is charged to naphtha stabilizer DA-203), JP-8 (which is sent to product storage after cooling), 550°F+ oil (which is recycled to reactor DC-201) and fuel gas (which flows to the boiler in the SNG plant).

- Unstabilized naphtha from DA-201 is combined with unstabilized naphtha from area 100 and charged to naphtha stabilizer DA-203 after being preheated in exchanger EA-205. Heat for reboiling the naphtha stabilizer is obtained by heat exchange with the hot jet fuel product.

- Fuel gas from the naphtha stabilizer is combined with fuel gas from FA-206 and is routed to the SNG boilers.

- The stabilized naphtha is cooled and sent to storage. It is also sent to the SNG boilers to be used as fuel.

Table 2.2 Hydrocracker Conditions

Reactor Conditions	5 Fixed Beds
Catalyst, % of Total & Type	
Bed 1	10%, HDS/HDN/HC
Bed 2	22.5%, HDS/HDN/HC
Beds 3-5	22.5% HC
WHSV, hr^{-1}	1.5
Average Reactor Temp.	675°F
Temperature Increase	
Bed 1	15°F
Bed 2-5	40°F
Heat of Reaction	20,000 BTU/#Mole H2
Reactor Pressure	
Inlet	1200 psig
Outlet	1175 psig
Recycle Rate H2	13,000 scf/Bbl
Conversion/Pass	50%
Catalyst Replacement	3 years @ $6/#

2.3 PSA Hydrogen Unit & Recompression (Area 300)

2.3.1 Hydrogen for both the hydrotreator and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig	
Temp.	65 ^{o}F	
Composition	mol%	
H2		63.19
CO		18.61
CO2		1.48
CH4		16.21
C2H6		0.31
COS, H2S, CS2		< 0.01
N2 + Ar		0.19
H2O		< 0.01

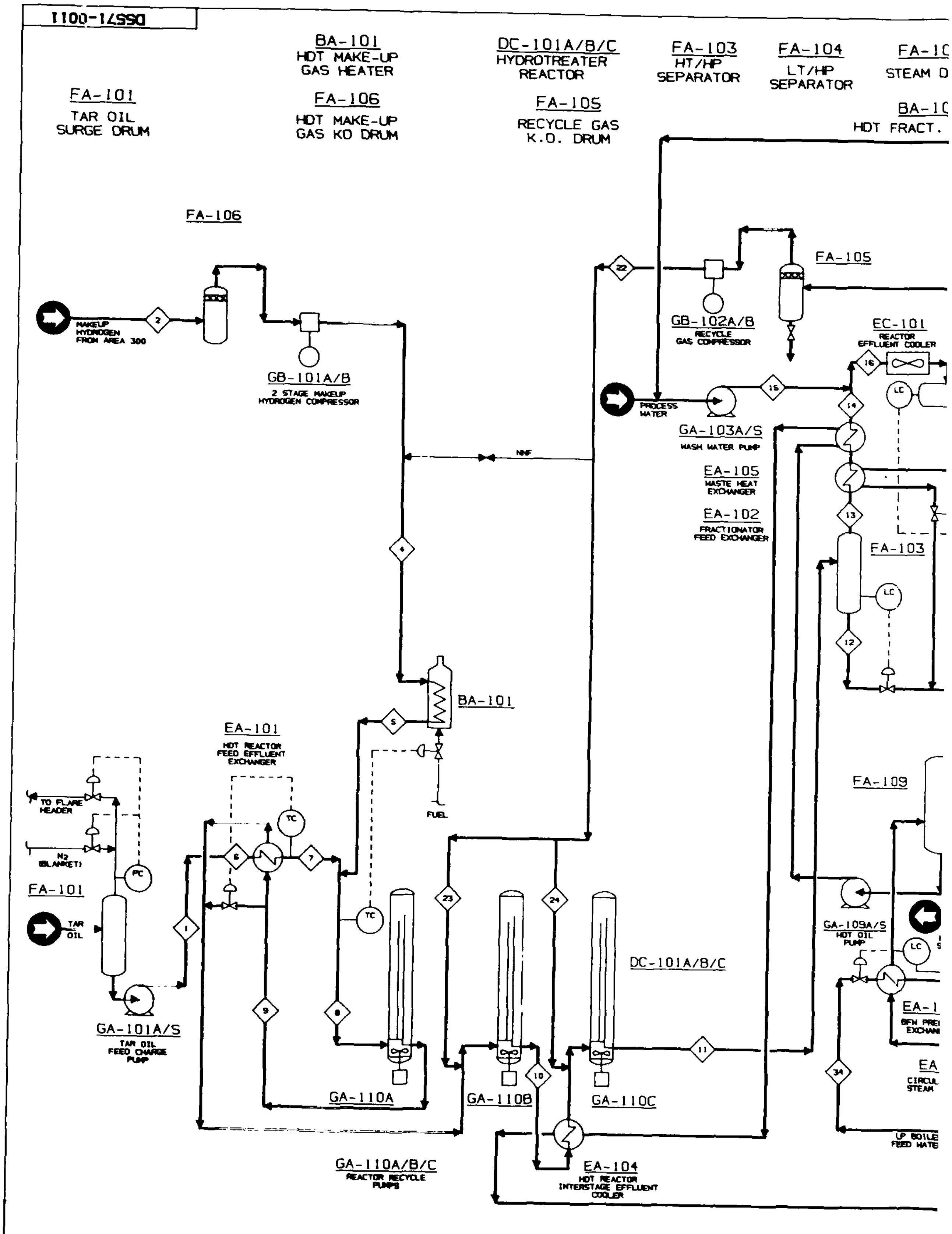
The PSA unit selectively absorbs all components expect H2 and produces a 99.99% vol. purity stream at about 345 psig and 80^{o}F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100^{o}F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

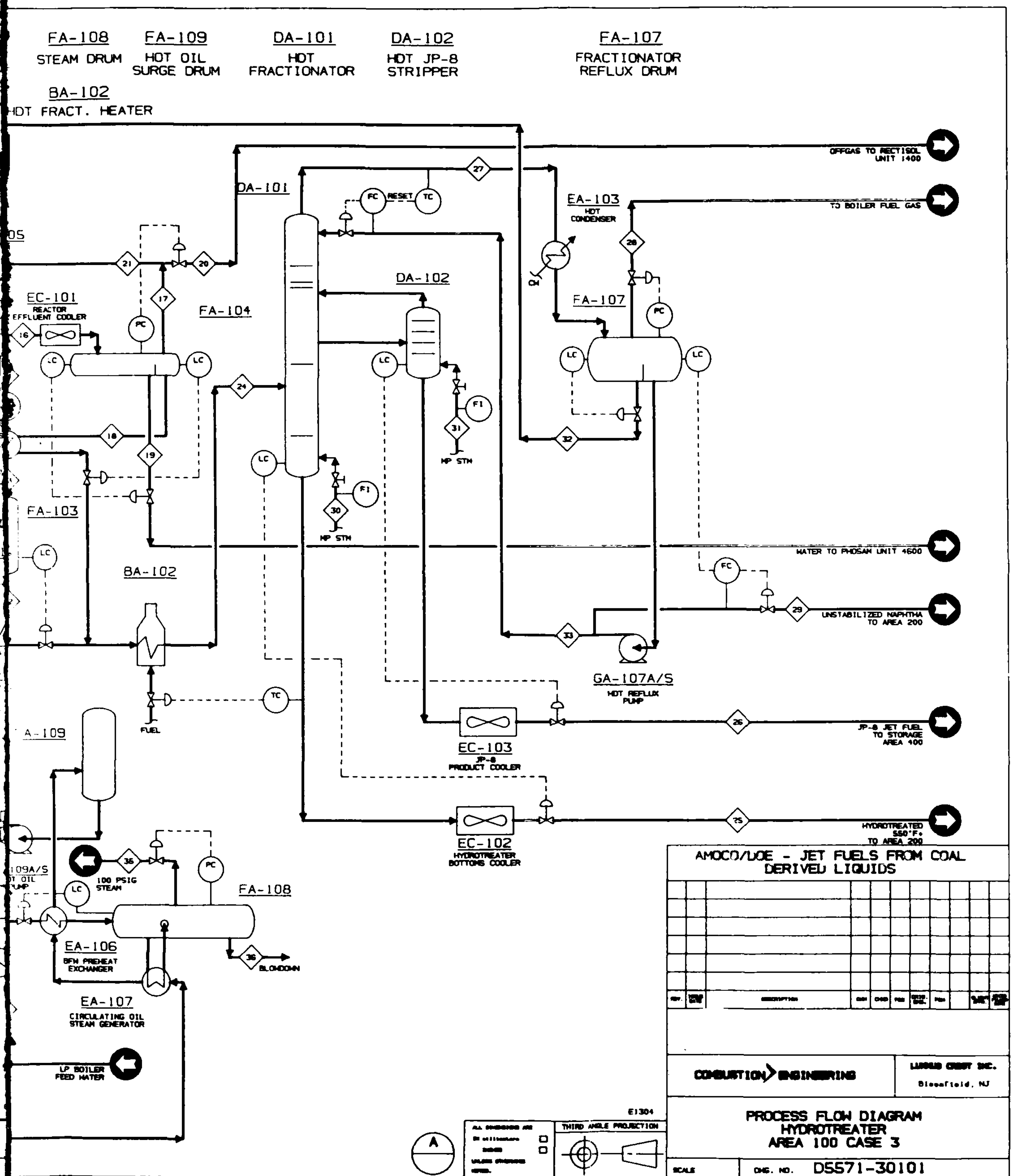
At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

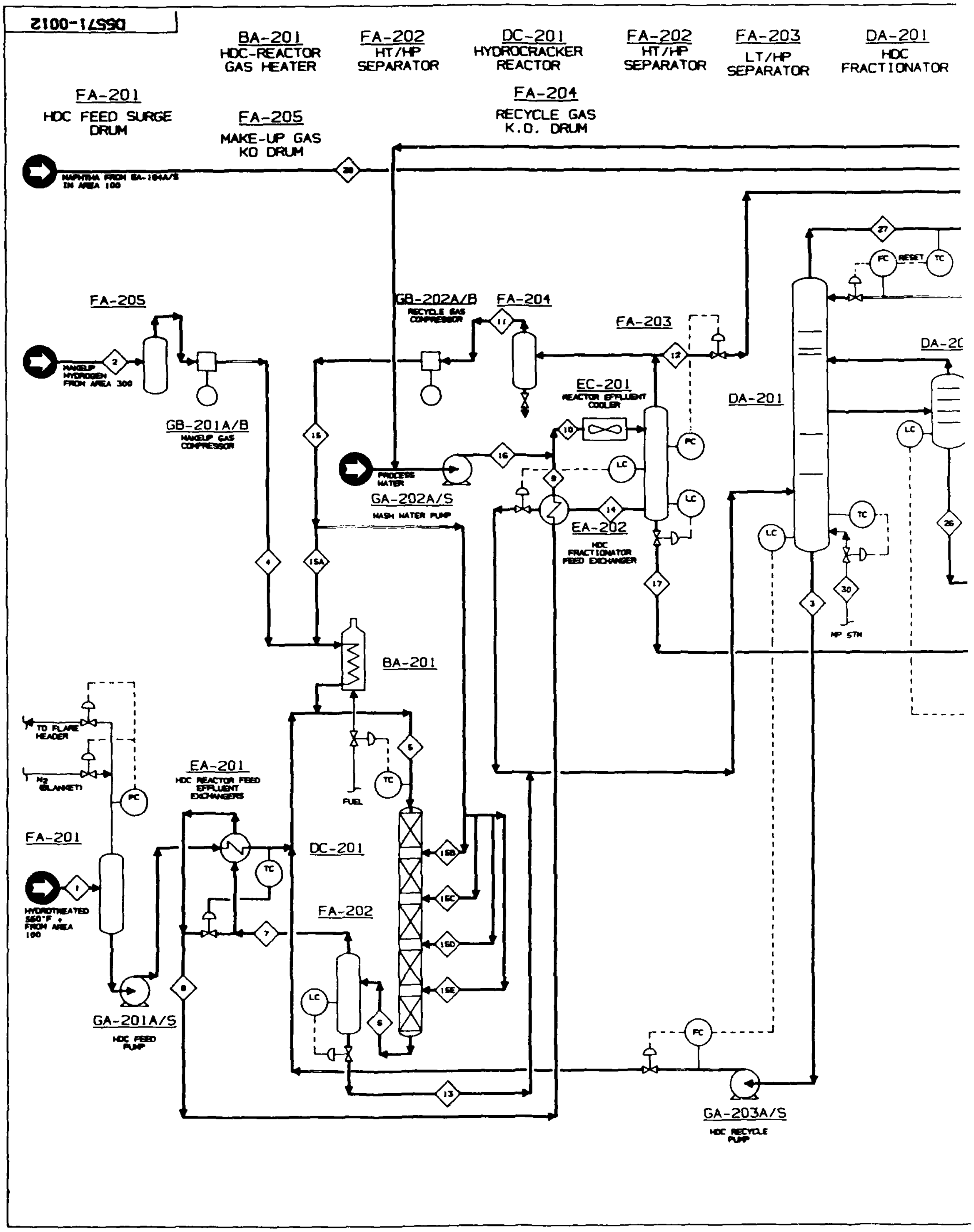
The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-30301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

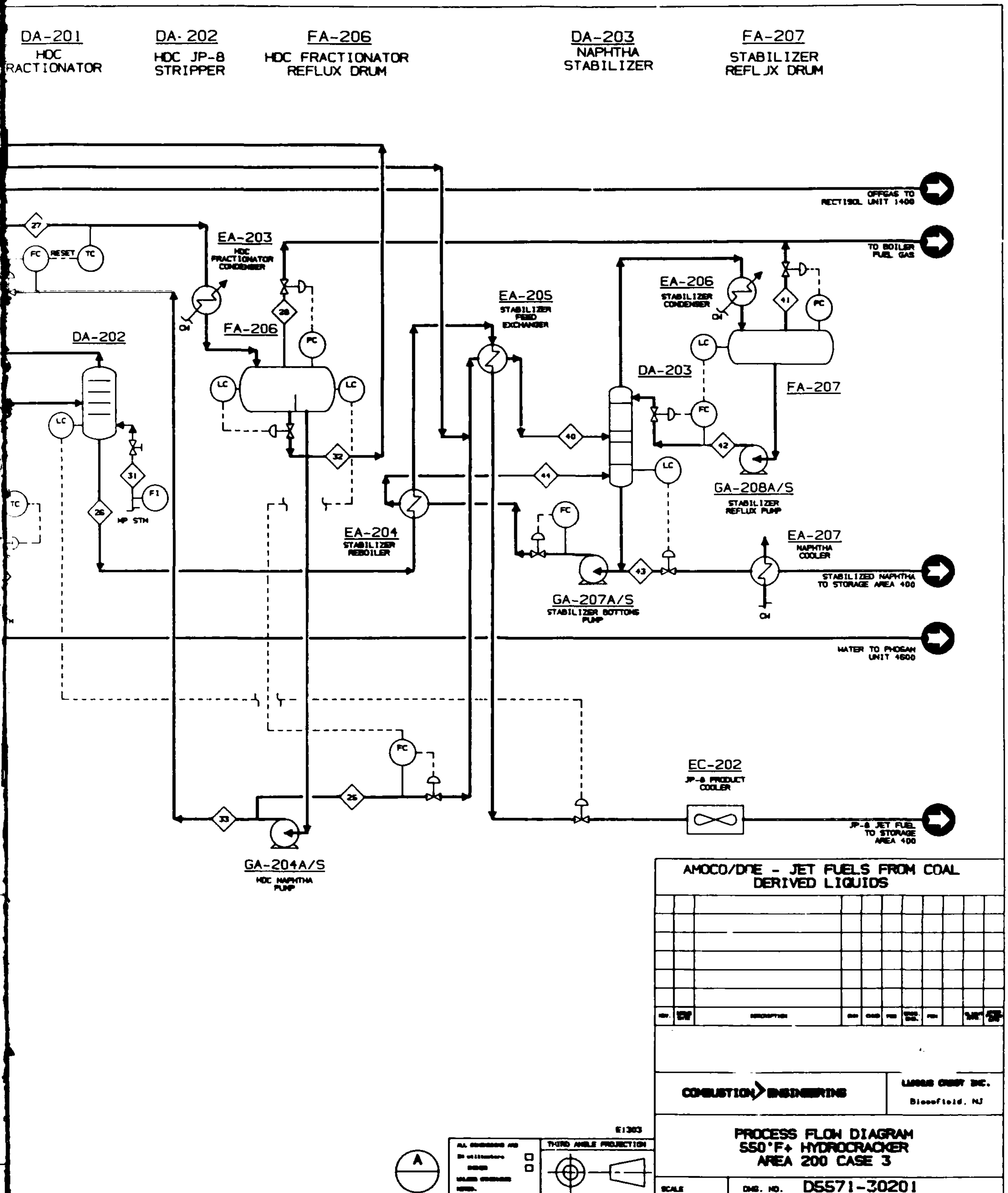


DSS71-0011
FA-101
TAR OIL
SURGE DRUM
BA-101
HDT MAKE-UP
GAS HEATER
FA-106
HDT MAKE-UP
GAS KO DRUM
DC-101A/B/C
HYDROTREATER
REACTOR
FA-105
RECYCLE GAS
K.O. DRUM
FA-103
HT/HP
SEPARATOR
FA-104
LT/HP
SEPARATOR
STEAM D
HDT FRACT.
FA-106
MAKEUP
HYDROGEN
FROM AREA 300
GB-101A/B
2 STAGE MAKEUP
HYDROGEN COMPRESSOR
GB-102A/B
RECYCLE
GAS COMPRESSOR
FA-105
EC-101
REACTOR
EFFLUENT COOLER
PROCESS
WATER
GA-103A/S
WASH WATER PUMP
EA-105
WASTE HEAT
EXCHANGER
EA-102
FRACTIONATOR
FEED EXCHANGER
FA-103
NNF
BA-101
FUEL
EA-101
HDT REACTOR
FEED EFFLUENT
EXCHANGER
TO FLARE
HEADER
N2
(BLANKET)
FA-101
TAR
OIL
GA-101A/S
TAR OIL
FEED CHARGE
PUMP
GA-110A
GA-110B
GA-110C
DC-101A/B/C
GA-110A/B/C
REACTOR RECYCLE
PUMPS
EA-104
HDT REACTOR
INTERSTAGE EFFLUENT
COOLER
FA-109
GA-109A/S
HDT OIL
PUMP
LP BOILER
FEED WATER


FA-108
STEAM DRUM
FA-109
HOT OIL SURGE DRUM
DA-101
HDT FRACTIONATOR
DA-102
HDT JP-8 STRIPPER
FA-107
FRACTIONATOR REFLUX DRUM
BA-102
HDT FRACT. HEATER
OFFGAS TO RECTISOL UNIT 1400
TO BOILER FUEL GAS
EA-103
HDT CONDENSER
RESET
EC-101
REACTOR EFFLUENT COOLER
FA-104
FA-103
HP STM
WATER TO PHOSAM UNIT 4600
UNSTABILIZED NAPHTHA TO AREA 200
GA-107A/S
HDT REFLUX PUMP
FUEL
JP-8 JET FUEL TO STORAGE AREA 400
EC-103
JP-8 PRODUCT COOLER
HYDROTREATED 550°F+ TO AREA 200
EC-102
HYDROTREATER BOTTOMS COOLER
100 PSIG STEAM
EA-106
BFW PREHEAT EXCHANGER
BLOWDOWN
EA-107
CIRCULATING OIL STEAM GENERATOR
LP BOILER FEED WATER
AMOCO/DOE - JET FUELS FROM COAL DERIVED LIQUIDS
COMBUSTION ENGINEERING
Bloomfield, NJ
THIRD ANGLE PROJECTION
E1304
PROCESS FLOW DIAGRAM
HYDROTREATER
AREA 100 CASE 3
SCALE
DWG. NO. D5571-30101

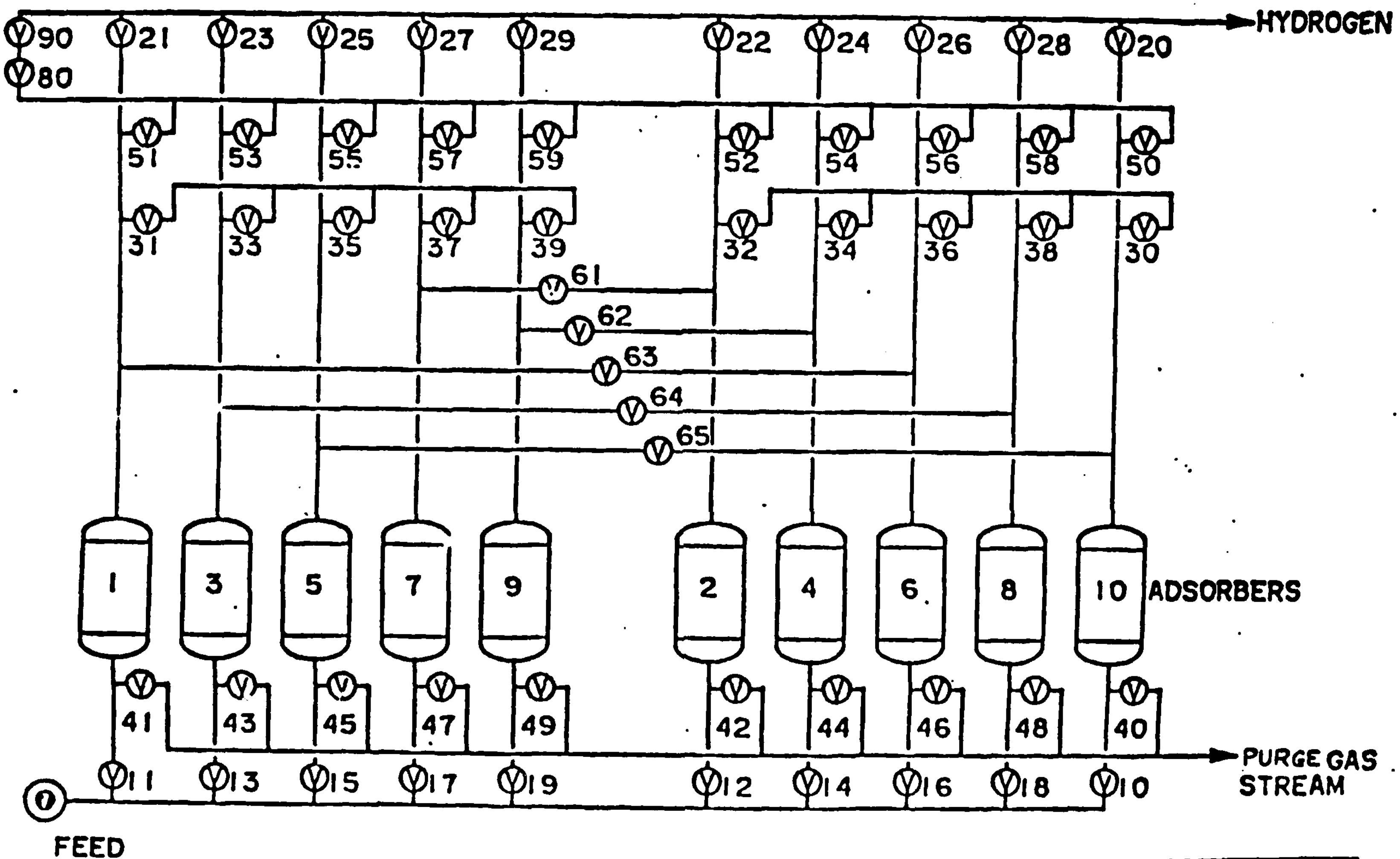


BA-201
HDC-REACTOR GAS HEATER
FA-202
HT/HP SEPARATOR
DC-201
HYDROCRACKER REACTOR
FA-202
HT/HP SEPARATOR
FA-203
LT/HP SEPARATOR
DA-201
HDC FRACTIONATOR
FA-201
HDC FEED SURGE DRUM
FA-205
MAKE-UP GAS KO DRUM
FA-204
RECYCLE GAS K.O. DRUM
NAPHTHA FROM GA-104A/S IN AREA 100
FA-205
MAKEUP HYDROGEN FROM AREA 300
GB-201A/B
MAKEUP GAS COMPRESSOR
GB-202A/B
RECYCLE GAS COMPRESSOR
FA-204
FA-203
EC-201
REACTOR EFFLUENT COOLER
DA-201
PROCESS WATER
GA-202A/S
WASH WATER PUMP
EA-202
HDC FRACTIONATOR FEED EXCHANGER
HP STM
BA-201
TO FLARE HEADER
N2 (BLANKET)
EA-201
HDC REACTOR FEED EFFLUENT EXCHANGERS
FUEL
FA-201
DC-201
HYDROTREATED 560°F + FROM AREA 100
FA-202
GA-201A/S
HDC FEED PUMP
GA-203A/S
HDC RECYCLE PUMP



DA-201
HDC
FRACTIONATOR
DA-202
HDC JP-8
STRIPPER
FA-206
HDC FRACTIONATOR
REFLUX DRUM
DA-203
NAPHTHA
STABILIZER
FA-207
STABILIZER
REFLUX DRUM
OFFGAS TO
RECTISOL UNIT 1400
TO BOILER
FUEL GAS
EA-203
HDC
FRACTIONATOR
CONDENSER
FC
RESET
TC
CW
FA-206
PC
LC
DA-202
EA-205
STABILIZER
FEED
EXCHANGER
EA-206
STABILIZER
CONDENSER
DA-203
FA-207
HP STM
FI
EA-204
STABILIZER
REBOILER
GA-208A/S
STABILIZER
REFLUX PUMP
EA-207
NAPHTHA
COOLER
STABILIZED NAPHTHA
TO STORAGE AREA 400
GA-207A/S
STABILIZER BOTTOMS
PUMP
WATER TO PHOSAM
UNIT 4600
EC-202
JP-8 PRODUCT
COOLER
JP-8 JET FUEL
TO STORAGE
AREA 400
GA-204A/S
HDC NAPHTHA
PUMP
AMOCO/DOE - JET FUELS FROM COAL
DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
PROCESS FLOW DIAGRAM
550°F+ HYDROCRACKER
AREA 200 CASE 3
E1303
THIRD ANGLE PROJECTION
SCALE
DWG. NO.
D5571-30201
A

SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



TYPICAL ARRANGEMENT
NUMBER OF ABSORBERS FOR THIS CASE = 10

LUMMUS	THE LUMMUS COMPANY
TITLE PSA HYDROGEN UNIT	
CLIENT AMOCO/DOE	AREA 300
PROJ NO 5571	
CASE 3	
1/3 FOR SUBTASK 1.2	ML E.S.
5571-...	

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 3 - MAXIMUM JP-8

AREA 100 -	HYDROTREATER
TAG NO. -	DESCRIPTION
BA-101	HDT Makeup Gas Heater
DA-101	HDT Fractionator
DA-102	JP-8 Stripper
DC-101A,B,C	Hydrotreater Reactors
EA-101	HDT Reactor Feed/Effl. Exch.
EA-102	Fract. Feed Exch.
EA-103	HDT Condenser
EA-104	HDT Reactor Int. Stg. Clr.
EA-105	Waste Heat Exchanger
EA-106	BFW Preheat Exch.
EA-107	Circulating Oil Stm. Gen.
EC-101	Reactor Effl. Cooler
EC-102	Hydrotreater Btms. Cooler
EC-103	JP-8 Product Cooler
FA-101	Tar Oil Surge Drum
FA-102	HDT Makeup Gas KO Drum
FA-103	HT/HP Separator
FA-104	LT/HP Separator
FA-105	Recycle Gas KO Drum
FA-107	Fractionator Reflux Drum
FA-108	Steam Drum
FA-109	Hot Oil Surge Drum
GA-101A/S	Tar Oil Feed Charge Pump
GA-103A/S	Wash Water Pump
GA-107A/S	HDT Reflux Pump
GA-109A/S	Hot Oil Pump
GA-110A/B/C	Reactor Recycle Pump
GB-101A/B	H_2 Makeup Compr.
GB-102A/B	Recycle Gas Compr.

CASE 3 - MAXIMUM JP-8 - Cont'd

AREA 200 - HYDROCRACKER

TAG. NO.	DESCRIPTION
BA-201	HDC Reactor Gas Heater
DA-201	HDC Fractionator
DA-202	HDC JP-8 Stripper
DA-203	Naphtha Stabilizer
DC-201	HDC Reactor
EA-201	HDC Reactor Feed/Effl. Exch.
EA-202	HDC Fract. Feed Exch.
EA-203	HDC Fract. Condenser
EA-204	Stabilizer Reboiler
EA-205	Stabilizer Feed Exch.
EA-206	Stabilizer Condenser
EA-207	Naphtha Cooler
EC-201	Reactor Effl. Cooler
EC-202	JP-8 Product Cooler
FA-201	HDC Feed Surge Drum
FA-202	HT/HP Separator
FA-203	LT/HP Separator
FA-204	Recycle Gas KO Drum
FA-205	Makeup Gas KO Drum
FA-206	HDC Fract. Reflux Drum
FA-207	Stabilizer Reflux Drum
FA-208	Fuel Oil Day Tank
GA-201A/S	HDC Feed Pump
GA-202A/S	Wash Water Pump
GA-203A/S	HDC Recycle Pump
GA-204A/S	HDC Naphtha Pump
GA-207A/S	Stabilizer Btms Pump
GA-208A/S	Stabilizer Reflux Pump
GA-209A/S	Fuel Oil Pump
GB-201A/B	Makeup Gas Compr.
GB-202A/B	Recycle Gas Compr.

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

W16214/bf

CASE 3 - MAXIMUM JP-8 - Cont'd

TAG NO.	DESCRIPTION
AREA 400 -	STORAGE AREA
FB-401	Jet Fuel Storage Tank
FB-402	Naphtha Storage Tank
FB-403	Fuel Oil Storage Tank
GA-401A/S	Tar/Tar Oil Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump
AREA 500 -	CATALYST HANDLING
TAG NO.	DESCRIPTION
FA-501	Catalyst Oil Drum
FA-502	Catalyst Storage Hopper
FA-503	Catalyst Transfer Vessels
FA-504	Spent Catalyst Vessel
FL-501	Catalyst Screen
GA-501A/S	Catalyst Transfer Pump
GA-502A/S	Catalyst Oil Pump

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for high pressure equipment purchased for various hydrotreater/hydrocracker projects. The unit pricing is some what conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the high pressure processing, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

	(Thousands of $) Case 3
Area 100 Hydrotreater	$20,702
Area 200 Hydrocracker	10,012
Area 300 PSA & Recompression	8,300
Area 400 OSBL	4,500
Area 500 Catalyst Handling	1,285
Total	$44,799

W16214/bf

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

Items	Equipment Type	$ Value	%. Comm	$ Comm.
1	Heaters	50	120	60
2	Towers	39	140	55
-	Internals	7	-	-
3	Reactors	1650	70	1155
13	Exchangers	582	70	407
3	Air Coolers	113	100	113
8	Vessels	388	85	330
11	Pumps	739	80	591
4	Compressors	1700	60	1020
	Special	-		
45	Total	$5268		$3731

Equipment	5268	
Commodities	3731	
Labor	2765	10% Equip. 60% Comm.
Indirects	2765	100%
Office	2723	45 pcs x 1100 x $55-
Subtotal	17,252	
Contingency	3,450	20%
Total	$20,702	

W16214/bf

3.2.3 Estimate Breakdown - Cont'd

Area 200

Items	Equipment Type	$ Valve	%. Comm	$ Comm.
1	Heaters	130	120	156
3	Towers	54	140	70
-	Internals	9	-	-
1	Reactors	325	85	276
8	Exchangers	188	100	188
2	Air Coolers	46	110	50
8	Vessels	258	100	258
14	Pumps	152	120	182
4	Compressors	550	100	550
	Special	-		
41	Total	$1712		$1730

Equipment	1712	
Commodities	1730	
Labor	1210	10% Equip. 60% Comm.
Indirects	1210	100%
Office	2481	41 pcs x 1100 x $55-
Subtotal	8343	
Contingency	1669	20%
Total	$10,012	

Area 300

PSA-unit 20 mm SCFD budget quote $3500

PSA unit 14 mm SCFD 2500

Installation 50% 1300

Subtotal $3800

Compressor	3400	1700 x 2.0 T.I.C.
Drum	300	100 x 3.0 T.I.C.
Subtotal	$7500	
Contingency	800	10%
Total	$8300	

W16214/bf

3.2.3 Estimate Breakdown - Cont'd

Area 400

Case 1 TIC = $5100

For Case 3 Deduct	Equipment	3 x 150 =	450
	Piping		200
	Total Deduct		$650

Case 3 Total - $4500

Area 500

Items	Equipment Type	$ Value	%. Comm	$ Comm.
4	Vessels	105	120	126
4	Pumps	48	120	58
8	Total	$153		$184

Equipment	153	
Commodities	184	
Labor	125	10% Equip. 60% Comm.
Indirects	125	100%
Office	484	8 pcs x 1100 x $55-
Subtotal	$1071	
Contingency	214	20%
	$1285	

W16214/bf

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require 7 men/shift to operate the plant broken down as follows:

Foreman	1	
Control Room	1	
HDT Operator	2	
HCR Operator	2	
PSA & relief man	1	
	7	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 5 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	7 positions x 4 people/position -	28
Supervisor & Admin.		5
QC Technician		1
Maintenance		5
Other (Stores or Janitorial)		1
Total		40

4.2 Utilities

The following utilities have been estimated from the computer simulations:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	1978 BPSD	$16/Bbl [(a)]	31648
SNG equivalent of Syn Gas & Purge Gas	3.7 MMSCFD	$3.80/MM Btu [(b)]	13780
Cooling Water	2040 GPM	$0.155/MGal [(c)]	456
Power	5450 kW	$0.04/kWH [(c)]	5232
Process Water	20.5 GPM	$0.45/MGal [(c)]	13

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

Catalyst	Use	Cost	$/SD
HDT Cat.	0.30 #/Bbl	$3.00/#	2864
HCR Cat.	0.0095 #/Bbl	$6.00/#	47
Inhibitors	50 PPM	$10/Gal	52
			2963

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

$$0.00005 \times 44{,}799{,}000 = \$2240./SD$$

W16214/bf

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-8 are proposed to be located to the east of the Rectisol Unit and Main Control Room of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 300' x 200' will be surrounded by an access road and will be divided by a central east-west road. Areas 100 & 500 will be located to the north and Areas 200 & 300 to the south.

A diked storage tank area approx. 360' x 265' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

Approximately 2000 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

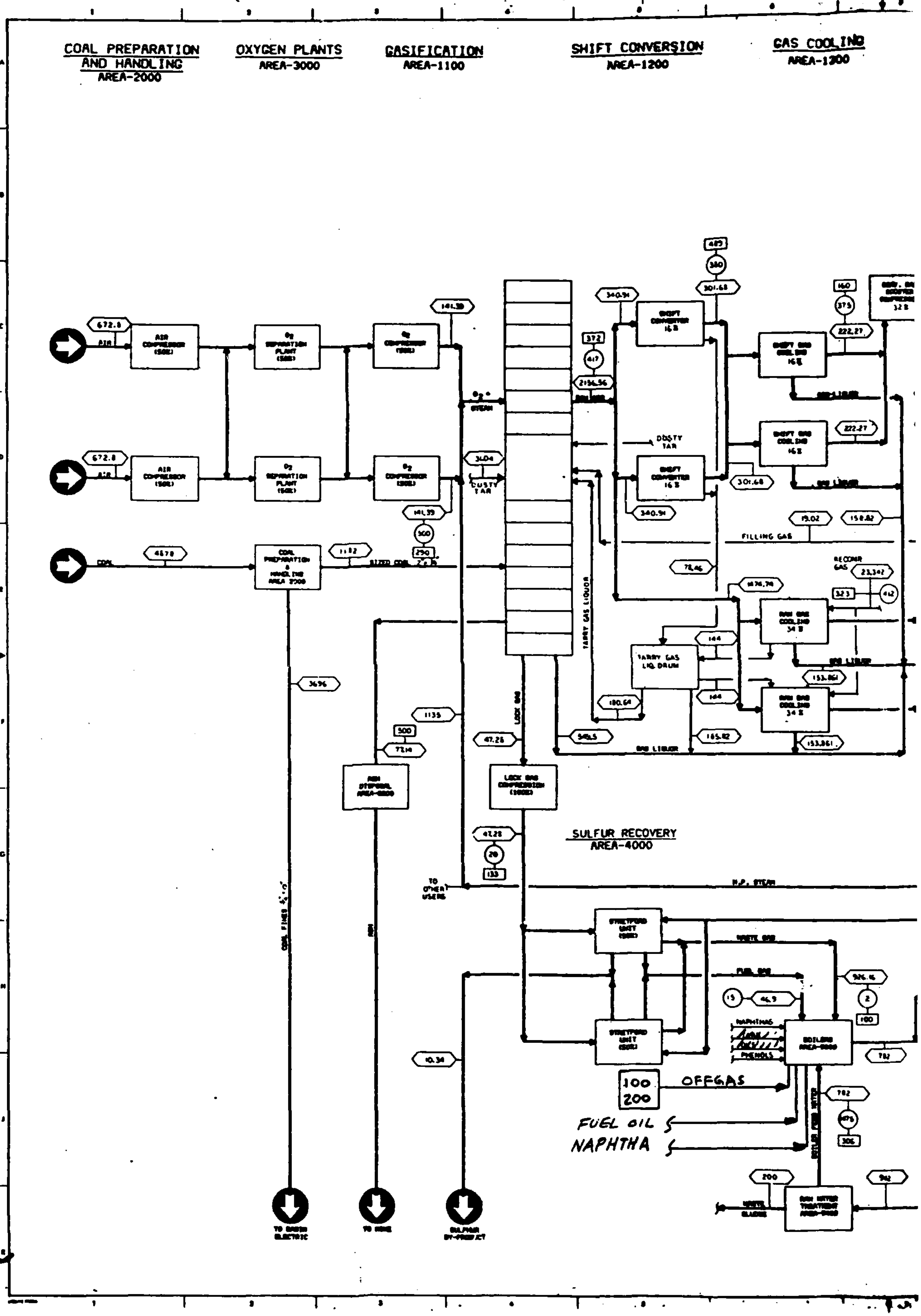
New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

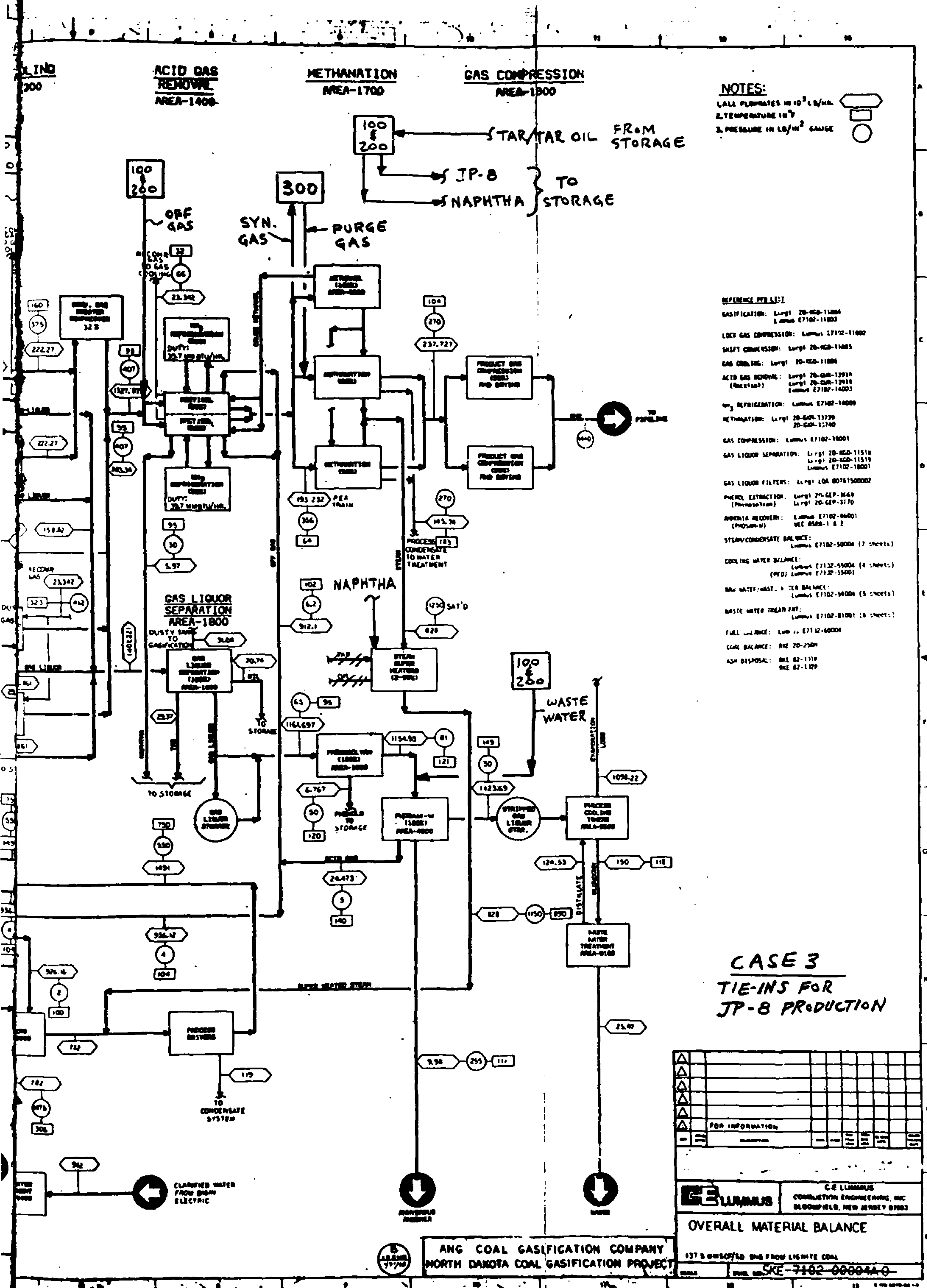
A summary of the lines is shown in table 5.1.

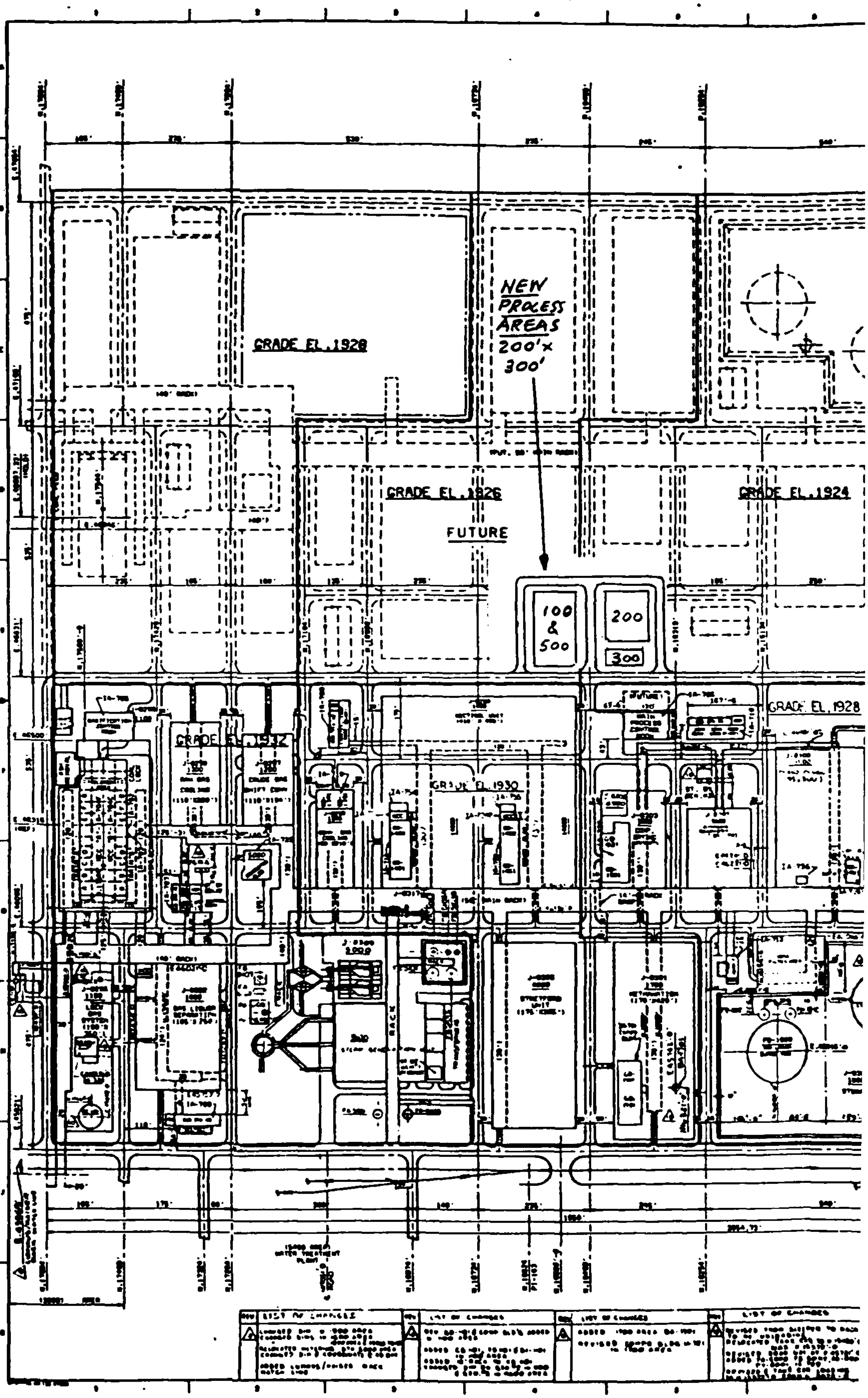
TABLE 5.1

INTERCONNECTING PIPING

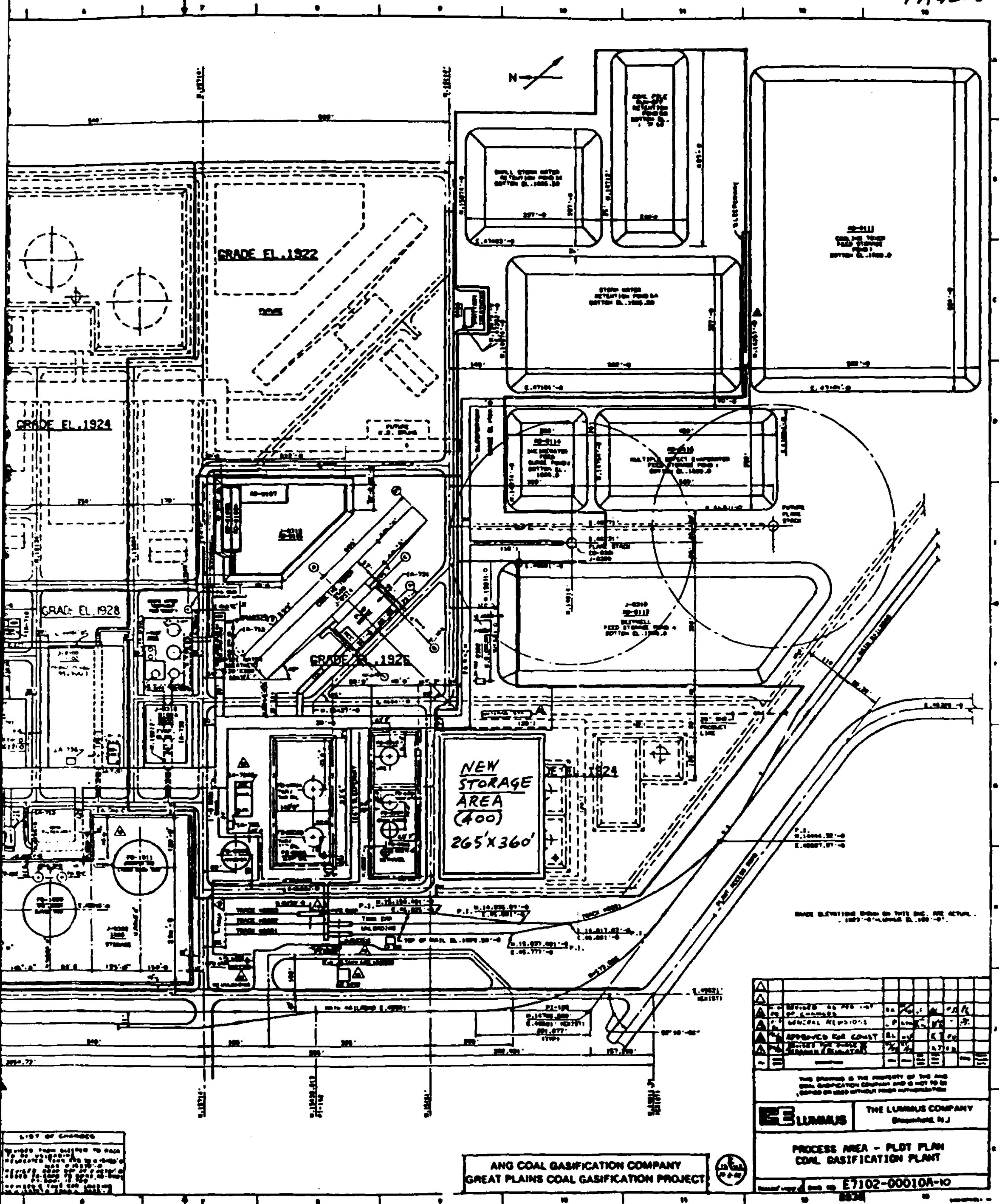
EST. SIZE	SERVICE	TO/FROM
4"	Tar/Tar Oil (Elec. Tr.)	Storage
3"	JP-8 Product	Storage
2"	Naphtha Product	Storage
16"	Wet Flare (Trace)	Flare
8"	Synthesis Gas	PSA/Rectisol
6"	Purge Gas	Methanation/PSA
1 1/2"	Off Gas	Rectisol/HDT,HDC
2"	Nitrogen	Main Rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
6"	M.P. Steam	"
1 1/2	Stm Cond.	"
1 1/2	BFW	"
1 1/2	Boiler B.D.	"
12"	C. W. Supply & Return	"
2"	Waste Water	Phosam/HDT,HDC
4"	Fuel Oil	Exist TKS/New TKS.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers


COAL PREPARATION AND HANDLING
AREA-2000
OXYGEN PLANTS
AREA-3000
GASIFICATION
AREA-1100
SHIFT CONVERSION
AREA-1200
GAS COOLING
AREA-1300
SULFUR RECOVERY
AREA-4000
AIR
COAL
672.8
4678
AIR COMPRESSOR
COAL PREPARATION & HANDLING AREA 2000
SIZED COAL
DUSTY TAR
TARRY GAS LIQUOR
TARRY GAS LIQ. DRUM
FILLING GAS
GAS LIQUOR
LOCK GAS
H.P. STEAM
TO OTHER USERS
WASTE GAS
FUEL GAS
NAPHTHAS
PHENOLS
100
200
OFFGAS
FUEL OIL
NAPHTHA
COAL FINES
ASH


ACID GAS REMOVAL
AREA-1400
METHANATION
AREA-1700
GAS COMPRESSION
AREA-1900
NOTES:
1. ALL FLOWRATES IN 10^3 LB/HR.
2. TEMPERATURE IN °F
3. PRESSURE IN LB/IN² GAUGE
100 & 200
TAR/TAR OIL FROM STORAGE
JP-8
NAPHTHA
TO STORAGE
100 200
300
OFF GAS
SYN. GAS
PURGE GAS
GAS LIQUOR SEPARATION
AREA-1800
NAPHTHA
100 & 200
WASTE WATER
TO PIPELINE
TO STORAGE
SUPER HEATED STEAM
PROCESS DRIVERS
TO CONDENSATE SYSTEM
CLARIFIED WATER FROM BASIN ELECTRIC
CASE 3
TIE-INS FOR JP-8 PRODUCTION
FOR INFORMATION
C-E LUMMUS
COMBUSTION ENGINEERING, INC.
BLOOMFIELD, NEW JERSEY 07003
OVERALL MATERIAL BALANCE
ANG COAL GASIFICATION COMPANY
NORTH DAKOTA COAL GASIFICATION PROJECT
SKE-7102-00004A-0



NEW PROCESS AREAS 200'x 300'
GRADE EL.1928
GRADE EL.1926
GRADE EL.1924
FUTURE
100 & 500
200
300
GRADE EL.1932
GRADE EL.1930
LIST OF CHANGES


GRADE EL. 1922
GRADE EL. 1924
GRADE EL. 1928
GRADE EL. 1926
NEW STORAGE AREA (400) 265'X360'
THE LUMMUS COMPANY
LUMMUS
PROCESS AREA - PLOT PLAN
COAL GASIFICATION PLANT
E7102-00010A-10
ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT

CASE 3

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PROBLEM C3U100 SOLUTION NOV 17 1987

PAGE A-1
CASE 3
AREA 100

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	1	2	3	4	23	24
STREAM NAME.........	FEED TAR	MAKEUP H2	HOT TAR OIL	NET MAKEUP	101B REC GAS	101C REC GAS
STREAM PHASE........	LIQUID	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F	154.0000	70.0000	550.0000	395.9943	151.1667	151.1667
RATE LB MOLS/HR	304.8536	1450.0000	304.8535	1449.9998	728.4996	672.4612
RATE LB /HR	47707.5703	2925.2324	47707.5703	2925.2324	2488.5923	2297.1611
ENTHALPY MM BTU /HR	2.0518	-3.0145	11.3451	0.1755	-0.9060	-0.8363
ENTHALPY BTU /LB	43.0074	-1030.5195	237.8040	59.9865	-364.0683	-364.0686
MOLECULAR WEIGHT....	156.4934	2.0174	156.4934	2.0174	3.4161	3.4160
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	2925.2324	0.0000	2925.2324	2488.5923	2297.1611
STD.RATE MM FT3/DAY	0.00	13.21	0.00	13.21	6.63	6.12
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4460	2.1428	2.1428
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	3.4161	3.4160
*** LIQUID PHASE ***						
RATE LB /HR	47707.5703	0.0000	47707.5703	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY	3269.42	0.00	3824.95	0.00	0.00	0.00
STD. LV RATE BBL/HR	133.10	0.00	133.10	0.00	0.00	0.00
CP, BTU /LB F	0.4045	0.0000	0.5913	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	156.4934	0.0000	156.4934	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	62.3749	0.0000	53.3155	0.0000	0.0000	0.0000
STD. API GRAVITY....	6.5896	0.0000	6.5896	0.0000	0.0000	0.0000
*** DRY BASIS ***						
RATE LB /HR	46753.7656	0.0000	46753.7656	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	185.5983	0.0000	185.5983	0.0000	0.0000	0.0000
UOP K	9.9530	0.0000	9.9530	0.0000	0.0000	0.0000
FLASH POINT, DEG F	142.3663	0.0000	142.3663	0.0000	0.0000	0.0000
CRIT. TEMP, F	966.2936	0.0000	966.2936	0.0000	0.0000	0.0000
CRIT. PRES, PSIA	424.8965	0.0000	424.8965	0.0000	0.0000	0.0000
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	2925.2324	0.0000	2925.2324	2477.1152	2286.5669
STD.RATE MM FT3/HR	0.00	0.55	0.00	0.55	0.28	0.25
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4460	2.1505	2.1506
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	3.4033	3.4033
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	46753.7656	0.0000	46753.7656	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY	3202.71	0.00	3736.27	0.00	0.00	0.00
CP, BTU /LB F	0.3926	0.0000	0.5720	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	185.5983	0.0000	185.5983	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	[illegible]	0.0000	53.4297	0.0000	0.0000	0.0000
STD. API GRAVITY....	[illegible]	[illegible]	6.5197	0.0000	0.0000	0.0000

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PROJECT GP JET FUELS ML
PROBLEM C3U10U SOLUTION NOV 17 1987

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	8	9	10	11	12	13
STREAM NAME..........	101A INLET	101A OUTLET	101B OUTLET	101C OUTLET	HOT SEP LIQ	HOT SEP VAP
STREAM PHASE........	MIXED	MIXED	MIXED	MIXED	LIQUID	VAPOR
TEMPERATURE, DEG F	467.2208	700.0000	700.0000	700.0000	699.9308	699.9308
RATE LB MOLS/HR	1754.8521	1363.4368	1714.6738	2037.5063	58.9225	1978.5837
RATE LB /HR	50632.8047	50633.4766	53122.7031	55423.3672	8365.1719	47058.1875
ENTHALPY MM BTU /HR	11.5204	22.5462	26.3515	29.9376	2.9036	27.0339
ENTHALPY BTU /LB	227.5274	445.2832	496.0495	540.1617	347.1082	574.4777
MOLECULAR WEIGHT....	28.8530	37.1357	30.9812	27.2016	141.9691	23.7838
*** VAPOR PHASE ***						
RATE LB /HR	5532.7441	18865.8438	32405.3750	47019.6406	C.0000	47058.1875
STD.RATE MM FT3/DAY	13.41	10.44	14.31	18.02	0.00	18.02
CP, BTU /LB F	1.9962	0.8624	0.8264	0.8150	0.0000	0.8149
MOLECULAR WEIGHT....	3.7569	16.4646	20.6213	23.7685	C.0000	23.7838
*** LIQUID PHASE ***						
RATE LB /HR	45100.0469	31767.6367	20717.3203	8403.7207	8365.1719	0.0000
ACT.RATE BBL/DAY	3645.26	3221.31	2161.44	901.53	897.70	0.00
STD. LV RATE BBL/HR	128.51	95.52	64.67	27.26	27.13	0.00
CP, BTU /LB F	0.5512	0.6658	0.6971	0.7214	C.7212	0.0000
MOLECULAR WEIGHT....	159.8356	145.9953	144.6516	141.7721	141.9691	0.0000
ACT.DENS LB /FT3	52.8860	42.1546	40.9716	39.8458	39.8324	0.0000
STD. API GRAVITY....	9.5333	17.3274	23.0012	29.0798	29.0353	0.0000
*** DRY BASIS ***						
RATE LB /HR	44961.8359	31526.7969	20533.1992	8322.6690	8294.7793	0.0000
MOLECULAR WEIGHT....	163.7996	154.3732	154.3828	151.9371	152.1266	0.0000
UOP K	10.0351	10.4339	10.7379	11.0075	11.0063	0.0000
FLASH POINT, DEG F	10.7020	-1.0973	-19.3667	-32.2335	-31.6788	0.0000
CRIT. TEMP, F	788.6456	660.3578	623.0140	579.9640	580.9435	0.0000
CRIT. PRES, PSIA	366.2994	348.6959	339.0001	329.3538	329.3210	0.0000
*** VAPOR PHASE ***						
RATE LB /HR	4717.1543	16974.6484	29310.3320	42758.4375	C.0000	42796.3281
STD.RATE MM FT3/HR	0.54	0.39	0.53	0.66	0.00	0.66
CP, BTU /LB F	2.2542	0.9020	0.8607	0.8457	C.0000	0.8455
MOLECULAR WEIGHT....	3.3047	16.3032	20.9412	24.5499	C.0000	24.5672
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	44961.8359	31526.7969	20533.1992	8322.6690	8294.7793	0.0000
ACT.RATE BBL/DAY	3633.59	3133.09	2132.21	898.67	884.95	0.00
CP, BTU /LB F	C.5492	0.6450	0.6718	0.6941	C.6941	0.0000
MOLECULAR WEIGHT....	163.7995	154.3732	154.3328	151.9371	152.1266	0.0000
ACT.DENS LB /FT3	52.6913	42.3375	41.1641	40.0328	40.0177	0.0000
STD. API GRAVITY....	9.5317	17.3159	23.1172	29.2650	29.2194	0.0000
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

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PROJECT GP JET FUELS ML
PROBLEM C3U100 SOLUTION NOV 17 1987

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	15	16	17	18	19	20
STREAM NAME..........	H2O WASH EFF	+ H2O	COLD SEP VAP	COLD SEP LIQ	SOUR H2O	PURGE GAS
STREAM PHASE........			VAPOR	LIQUID	LIQUID	VAPOR
TEMPERATURE, DEG F			120.0000	120.0000		120.0000
RATE LB MOLS/HR			1416.9827	303.4501		16.0176
RATE LB /HR			4840.5156	37544.5781		54.7174
ENTHALPY MM BTU /HR			-2.0745	0.4793		-0.0235
ENTHALPY BTU /LB			-428.5753	12.7654		-428.5779
MOLECULAR WEIGHT....			3.4161	123.7257		3.4161
*** VAPOR PHASE ***						
RATE LB /HR			4840.5156	0.0000		54.7174
STD.RATE MM FT3/DAY			12.91	0.00		0.15
CP, BTU /LB F			2.1473	0.0000		2.1464
MOLECULAR WEIGHT....			3.4161	0.0000		3.4161
*** LIQUID PHASE ***						
RATE LB /HR			0.0000	37544.5781		0.0000
ACT.RATE BBL/DAY			0.00	3249.67		0.00
STD. LV RATE BBL/HR			0.00	133.21		0.00
CP, BTU /LB F			0.0000	0.4767		0.0000
MOLECULAR WEIGHT....			0.0000	123.7257		0.0000
ACT.DENS LB /FT3			0.0000	49.3855		0.0000
STD. API GRAVITY....			0.0000	44.1161		0.0000
*** DRY BASIS ***						
RATE LB /HR			0.0000	37533.6641		0.0000
MOLECULAR WEIGHT....			0.0000	123.9373		0.0000
UOP K			0.0000	11.5026		0.0000
FLASH POINT, DEG F			0.0000	-77.8037		0.0000
CRIT. TEMP, F			0.0000	564.1465		0.0000
CRIT. PRES, PSIA			0.0000	442.3537		0.0000
*** VAPOR PHASE ***						
RATE LB /HR			4813.1924	0.0000		54.4651
STD.RATE MM FT3/HR			0.54	0.00		0.01
CP, BTU /LB F			2.1543	0.0000		2.1543
MOLECULAR WEIGHT....			3.4033	0.0000		3.4033
VISCOSITY, CP			0.0000	0.0000		0.0000
*** LIQUID PHASE ***						
RATE LB /HR			0.0000	37533.6641		0.0000
ACT.RATE BBL/DAY			0.00	3248.92		0.00
CP, BTU /LB F			0.0000	0.4766		0.0000
MOLECULAR WEIGHT....			0.0000	123.9373		0.0000
ACT.DENS LB /FT3			0.0000	49.3927		0.0000
STD. API GRAVITY....			0.0000	44.1260		0.0000
VISCOSITY, CP			[illegible]	0.0000		0.0000

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	21	22
STREAM NAME.........	RECYCLE GAS	REC COMP DIS
STREAM PHASE........	VAPOR	VAPOR
TEMPERATURE, DEG F	120.0000	151.1567
RATE LB MOLS/HR	1400.9648	1400.9609
RATE LB /HR	4785.7979	4795.7539
ENTHALPY MM BTU /HR	-2.0511	-1.7423
ENTHALPY BTU /LB	-428.5756	-364.0634
MOLECULAR WEIGHT....	3.4161	3.4161
*** VAPOR PHASE ***		
RATE LB /HR	4785.7979	4795.7539
STD.RATE MM FT3/DAY	12.76	12.76
CP, BTU /LB F	2.1464	2.1446
MOLECULAR WEIGHT....	3.4161	3.4161
*** LIQUID PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE BBL/DAY	0.00	0.00
STD. LV RATE BBL/HR	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
STD. API GRAVITY....	0.0000	0.0000
*** DRY BASIS ***		
RATE LB /HR	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
UOP K	0.0000	0.0000
FLASH POINT, DEG F	0.0000	0.0000
CRIT. TEMP, F	0.0000	0.0000
CRIT. PRES, PSIA	0.0000	0.0000
*** VAPOR PHASE ***		
RATE LB /HR	4763.7275	4763.6526
STD.RATE MM FT3/HR	0.53	0.53
CP, BTU /LB F	2.1543	2.1523
MOLECULAR WEIGHT....	3.4033	3.4033
VISCOSITY, CP	0.0000	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE BBL/DAY	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
STD. API GRAVITY....	0.0000	0.0000

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SIMULATION SCIENCES INC.
PROJECT GP JET FUELS | ML
PROBLEM C3U100 | SOLUTION | NOV 20 1987

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.		25	26	27	28	29
STREAM NAME.........	TOT FRAC FD	HDC FEED	HCT JP-8	FRAC OVHD	HDT HC VAP	HDT NAPHTHA
STREAM PHASE........	MIXED	LIQUID	LIQUID	VAPOR	VAPOR	LIQUID
TEMPERATURE, DEG F	543.3046	490.5446	383.7116	305.1799	90.0000	90.0000
PRESSURE, PSIA	35.0000	57.0000	55.5000	50.0000	40.0000	40.0000
RATE LB MOLS/HR	362.3723	50.4040	153.0457	572.4960	65.8279	93.1744
RATE LB /HR	45909.7734	11660.0059	23950.0117	42544.3281	1111.1794	9150.1523
ENTHALPY MM BTU /HR	15.6194	1.9736	3.5358	12.3012	0.1937	-0.0266
ENTHALPY BTU /LB	340.2186	169.2655	147.6333	289.1396	174.3571	-2.8925
MOLECULAR WEIGHT....	126.6923	231.3305	156.4893	74.3138	16.8801	98.6338
*** VAPOR PHASE ***						
RATE LB /HR	41286.4531	0.0000	0.0000	42544.3281	1111.1794	0.0000
ACT.RATE FT3/SEC	28.26	0.00	0.00	25.00	2.68	0.00
STD.RATE MM FT3/DAY	3.12	0.00	0.00	5.21	0.60	0.00
CP, BTU /LB F	0.5866	0.0000	0.0000	0.5071	0.5981	0.0000
MOLECULAR WEIGHT....	120.6747	0.0000	0.0000	74.3138	16.8801	0.0000
ACT.DENS LB /FT3	0.4058	0.0000	0.0000	0.4728	0.1151	0.0000
COMPRESSIBILITY (Z).	0.9671	0.0000	0.0000	0.9576	0.9948	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	4623.3184	11660.0059	23950.0117	0.0000	0.0000	9190.1523
ACT.RATE BBL/DAY	413.59	992.65	2363.61	0.00	0.00	858.17
STD. LV RATE BBL/HR	13.79	34.45	92.28	0.00	0.00	35.11
CP, BTU /LB F	0.6056	0.5736	0.6041	0.0000	0.0000	0.4903
MOLECULAR WEIGHT....	228.4018	231.3308	156.4893	0.0000	0.0000	98.6338
ACT.DENS LB /FT3	47.7831	50.2105	43.3135	0.0000	0.0000	45.7763
STD. API GRAVITY....	16.0970	14.7462	35.5351	0.0000	0.0000	57.5816
*** DRY BASIS ***						
RATE LB /HR	4623.2461	11640.5879	23898.9688	0.0000	0.0000	9188.4551
MOLECULAR WEIGHT....	228.4440	235.9922	159.1016	0.0000	0.0000	98.7155
UOP K	10.6390	10.5887	11.4572	0.0000	0.0000	11.7568
FLASH POINT, DEG F	144.2612	228.3018	110.9470	0.0000	0.0000	-18.0632
CRIT. TEMP, F	957.3212	987.3505	744.5731	0.0000	0.0000	532.0510
CRIT. PRES, PSIA	312.7897	306.0754	362.9194	0.0000	0.0000	482.2778
*** VAPOR PHASE ***						
RATE LB /HR	41195.2188	0.0000	0.0000	40607.9063	1090.6182	0.0000
ACT.RATE FT3/SEC	27.83	0.00	0.00	20.12	2.64	0.00
STD.RATE MM FT3/HR	0.13	0.00	0.00	0.18	0.02	0.00
CP, BTU /LB F	0.5869	0.0000	0.0000	0.5117	0.6009	0.0000
MOLECULAR WEIGHT....	122.2172	0.0000	0.0000	87.3287	16.8600	0.0000
ACT.DENS LB /FT3	0.4112	0.0000	0.0000	0.5606	0.1149	0.0000
COMPRESSIBILITY (Z).	0.9666	0.0000	0.0000	0.9490	0.9947	0.0000
VISCOSITY, CP	0.0120	0.0000	0.0000	0.0100	0.0112	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	4623.2461	11640.5879	23898.9688	0.0000	0.0000	9188.4551
ACT.RATE BBL/DAY	413.58	990.97	2359.59	0.00	0.00	858.05
CP, BTU /LB F	0.6055	0.5726	0.6031	0.0000	0.0000	0.4902
MOLECULAR WEIGHT....	228.4440	235.9922	159.1016	0.0000	0.0000	98.7155
ACT.DENS LB /FT3	47.7832	50.2113	43.2947	0.0000	0.0000	45.7741
STD. API GRAVITY....	16.0971	14.7540	35.5960	0.0000	0.0000	57.5903

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	30	31	32	33
STREAM NAME..........	FRAC STR STM	JP-8 STP STM	SOUR H2O	REFLUX
STREAM PHASE.........	VAPOR	VAPOR	LIQUID	LIQUID
TEMPERATURE, DEG F	314.0779	311.6984	90.0000	90.0000
PRESSURE, PSIA	57.0000	55.5000	40.0000	40.0000
RATE LB MOLS/HR	48.2931	57.7297	105.9445	307.5491
RATE LB /HR	869.9995	1039.9995	1908.5906	30334.7578
ENTHALPY MM BTU /HR	1.0352	1.2375	0.1107	-0.0877
ENTHALPY BTU /LB	1189.5887	1189.6687	58.0156	-2.8925
MOLECULAR WEIGHT....	18.0150	18.0150	18.0150	98.6339
*** VAPOR PHASE ***				
RATE LB /HR	869.9995	1039.9995	0.0000	0.0000
ACT.RATE FT3/SEC	1.89	2.32	0.00	0.00
STD.RATE MM FT3/DAY	0.44	0.53	0.00	0.00
CP, BTU /LB F	0.5323	0.5313	0.0000	0.0000
MOLECULAR WEIGHT....	18.0150	18.0150	0.0000	0.0000
ACT.DENS LB /FT3	0.1273	0.1248	0.0000	0.0000
COMPRESSIBILITY (Z).	0.9673	0.9682	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	0.0000	0.0000	1908.5906	30334.7578
ACT.RATE BBL/DAY	0.00	0.00	131.37	2832.64
STD. LV RATE BBL/HR	0.00	0.00	5.46	115.89
CP, BTU /LB F	0.0000	0.0000	0.9978	0.4904
MOLECULAR WEIGHT....	0.0000	0.0000	18.0150	98.6339
ACT.DENS LB /FT3	0.0000	0.0000	62.1002	45.7763
STD. API GRAVITY....	0.0000	0.0000	10.0635	57.5815
*** DRY BASIS ***				
RATE LB /HR	0.0000	0.0000	0.0000	30329.1484
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	98.7156
UOP K	0.0000	0.0000	0.0000	11.7568
FLASH POINT, DEG F	0.0000	0.0000	0.0000	-18.0632
CRIT. TEMP, F	0.0000	0.0000	0.0000	532.0514
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	482.2779
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	0.0000	0.0000	0.0000	30329.1484
ACT.RATE BBL/DAY	0.00	0.00	0.00	2832.26
CP, BTU /LB F	0.0000	0.0000	0.0000	0.4903
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	98.7156
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	45.7741
STD. API GRAVITY....	0.0000	0.0000	0.0000	57.5907
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

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CASE 3
AREA 200

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	1	2	3	4	5	6
STREAM NAME..........	HDT 550+	MAKEUP H2	HDC RECYCLE	NET MAKEUP	RX INLET	RX EFFLUENT
STREAM PHASE........	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	MIXED
TEMPERATURE, DEG F	200.0000	70.0000	510.0000	302.5687	662.0000	683.7249
PRESSURE, PSIA	1240.0000	340.0000	1240.0000	1310.0000	1200.0000	1175.0000
RATE LB MOLS/HR	50.5695	250.0000	43.5297	250.0000	613.2278	1060.3828
RATE LB /HR	11659.3164	504.3504	11304.9629	504.3504	24256.2656	25780.8398
ENTHALPY MM BTU /HR	0.2596	-0.5197	2.2272	-0.1211	8.3117	12.0444
ENTHALPY BTU /LB	22.2675	-1030.5193	197.0121	-240.1482	342.6631	467.1856
MOLECULAR WEIGHT....	230.5699	2.0174	259.7073	2.0174	39.5551	24.3128
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	504.3504	0.0000	504.3504	4392.6396	18838.0352
ACT.RATE FT3/SEC	0.00	1.18	0.00	0.45	1.49	3.07
STD.RATE MM FT3/DAY	0.00	2.28	0.00	2.28	4.72	9.34
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4769	1.2568	0.9427
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	8.4755	18.3688
ACT.DENS LB /FT3	0.0000	0.1189	0.0000	0.3098	0.8177	1.7035
COMPRESSIBILITY (Z).	0.0000	1.0145	0.0000	1.0429	1.0334	1.0326
*** LIQUID PHASE ***						
RATE LB /HR	11659.8164	0.0000	11304.9629	0.0000	19863.6133	6942.8047
ACT.RATE BBL/DAY	863.65	0.00	965.10	0.00	1981.23	761.47
STD. LV RATE BBL/HR	34.51	0.00	34.40	0.00	60.60	22.06
CP, BTU /LB F	0.4324	0.0000	0.6123	0.0000	0.6722	0.6923
MOLECULAR WEIGHT....	230.5699	0.0000	259.7070	0.0000	209.1985	199.2751
ACT.DENS LB /FT3	57.7093	0.0000	50.0713	0.0000	42.8563	38.9739
STD. API GRAVITY....	14.9850	0.0000	19.0979	0.0000	19.5101	25.7646
*** DRY BASIS ***						
RATE LB /HR	11659.8164	0.0000	11304.9629	0.0000	19863.0898	6942.5146
MOLECULAR WEIGHT....	230.5699	0.0000	259.7070	0.0000	209.2572	199.3592
UOP K	10.5850	0.0000	10.9939	0.0000	10.8408	11.1666
FLASH POINT, DEG F	226.2674	0.0000	243.4616	0.0000	120.2164	18.8510
CRIT. TEMP, F	977.5262	0.0000	939.0624	0.0000	776.8339	721.3593
CRIT. PRES, PSIA	319.4649	0.0000	261.3383	0.0000	273.5729	271.2650
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	504.3504	0.0000	504.3504	4385.6611	18816.2578
ACT.RATE FT3/SEC	0.00	1.18	0.00	0.45	1.49	3.07
STD.RATE MM FT3/HR	0.00	0.09	0.00	0.09	0.20	0.39
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4769	1.2581	0.9433
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	8.4683	18.3693
ACT.DENS LB /FT3	0.0000	0.1189	0.0000	0.3098	0.8170	1.7035
COMPRESSIBILITY (Z).	0.0000	1.0145	0.0000	1.0429	1.0334	1.0327
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	11659.8164	0.0000	11304.9629	0.0000	19863.0898	6942.5146
ACT.RATE BBL/DAY	863.65	0.00	965.10	0.00	1981.17	761.43
CP, BTU /LB F	0.4324	0.0000	0.6123	0.0000	0.6722	0.6922
MOLECULAR WEIGHT....	230.5699	0.0000	259.7070	0.0000	209.2572	199.3592
ACT.DENS LB /FT3	57.7093	0.0000	50.0713	0.0000	42.8566	38.9742
STD. API GRAVITY....	[illegible]	[illegible]	19.0979	0.0000	19.5103	25.7652

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	7	8	9	10	11	12
STREAM NAME.........	HOT SEP VAP	EFF FROM E-2	E-202 OUTLET	COLD SEP FEE	RECYCLE GAS	PURGE GAS
STREAM PHASE........	VAPOR	MIXED	MIXED	MIXED	VAPOR	VAPOR
TEMPERATURE, DEG F	683.6332	584.8490	379.3525	365.2185	120.0000	120.0000
PRESSURE, PSIA	1170.0000	1145.0000	1120.0000	1110.0000	1095.0000	1095.0000
RATE LB MOLS/HR	1025.7039	1025.7036	1025.7039	1039.5811	791.5505	115.9933
RATE LB /HR	18862.0391	18562.0391	18862.0352	19112.0352	2315.1357	339.2587
ENTHALPY MM BTU /HR	9.7359	7.6068	3.5190	3.5429	-1.2087	-0.1771
ENTHALPY BTU /LB	516.1650	403.2874	186.5671	185.3755	-522.0969	-522.0966
MOLECULAR WEIGHT....	18.3894	18.3894	18.3894	18.3844	2.9248	2.9248
*** VAPOR PHASE ***						
RATE LB /HR	18862.0391	13840.3379	7340.8643	7205.7803	2315.1357	339.2587
ACT.RATE FT3/SEC	3.08	2.81	2.20	2.21	1.30	~~0.19~~
STD.RATE MM FT3/DAY	9.34	9.10	8.69	8.78	7.21	1.06
CP, BTU /LB F	0.9423	1.0164	1.2813	1.2866	2.4858	2.4858
MOLECULAR WEIGHT....	18.3894	13.8471	7.6972	7.4761	2.9248	2.9248
ACT.DENS LB /FT3	1.6985	1.3699	0.9251	0.9067	0.4932	0.4932
COMPRESSIBILITY (Z).	1.0325	1.0326	1.0350	1.0340	1.0439	1.0439
*** LIQUID PHASE ***						
RATE LB /HR	0.0000	5021.7061	11521.1738	11906.2520	0.0000	0.0000
ACT.RATE BBL/DAY	0.00	522.39	1139.93	1172.41	0.00	0.00
STD. LV RATE BBL/HR	0.00	16.30	40.01	41.50	0.00	0.00
CP, BTU /LB F	0.0000	0.6770	0.6182	0.6133	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	191.7433	160.0094	157.2109	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	41.0911	43.2028	43.4098	0.0000	0.0000
STD. API GRAVITY....	0.0000	29.1072	40.3978	41.0296	0.0000	0.0000
*** DRY BASIS ***						
RATE LB /HR	0.0000	5021.4668	11520.3184	11894.9570	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	191.8315	160.1033	158.3731	0.0000	0.0000
UOP K	0.0000	11.2991	11.6781	11.6975	0.0000	0.0000
FLASH POINT, DEG F	0.0000	0.3040	-38.4711	-39.4843	0.0000	0.0000
CRIT. TEMP, F	0.0000	710.1705	634.0975	631.0258	0.0000	0.0000
CRIT. PRES, PSIA	0.0000	281.3445	318.6971	321.7490	0.0000	0.0000
*** VAPOR PHASE ***						
RATE LB /HR	18840.2578	13518.7949	7319.9404	6945.2979	2293.0664	336.0247
ACT.RATE FT3/SEC	3.08	2.80	2.20	2.18	1.30	0.19
STD.RATE MM FT3/HR	0.39	0.38	0.36	0.36	0.30	0.04
CP, BTU /LB F	0.9429	1.0172	1.2835	1.3173	2.5054	2.5054
MOLECULAR WEIGHT....	18.3898	13.8421	7.6847	7.3156	2.9014	2.9014
ACT.DENS LB /FT3	1.6985	1.3693	0.9235	0.8867	0.4892	0.4892
COMPRESSIBILITY (Z).	1.0325	1.0326	1.0351	1.0346	1.0440	1.0440
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	0.0000	5021.4668	11520.3184	11894.9570	0.0000	0.0000
ACT.RATE BBL/DAY	0.00	522.37	1139.86	1171.53	0.00	0.00
CP, BTU /LB F	0.0000	0.6769	0.6182	0.6129	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	191.8315	160.1033	158.3731	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	41.0910	43.2032	43.4011	0.0000	0.0000
STD. API GRAVITY....	0.0000	[illegible]	[illegible]	41.0590	0.0000	0.0000
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	0.0000	0.0000

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	13	14	15	16	17
STREAM NAME..........	HOT SEP LIQ	COLD SEP LIQ	REC COMP DIS	H2O WASH	SOUR H2O
STREAM PHASE.........	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID
TEMPERATURE, DEG F	683.6332	120.0000	153.7265	125.0000	120.0000
PRESSURE, PSIA	1170.0000	1095.0000	1265.0000	1095.0000	1095.0000
RATE LB MOLS/HR	34.6793	118.5703	791.5386	13.8773	13.4675
RATE LB /HR	6918.8027	16215.1074	2315.1602	249.9999	242.5315
ENTHALPY MM BTU /HR	2.3081	0.2972	-1.0187	0.0239	0.0063
ENTHALPY BTU /LB	333.5910	18.3297	-439.9921	95.7615	25.9441
MOLECULAR WEIGHT....	199.5084	136.7552	2.9248	18.0150	18.0087
*** VAPOR PHASE ***					
RATE LB /HR	0.0000	0.0000	2315.1602	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	1.20	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00	7.21	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	2.4957	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	2.9248	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.5358	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	1.0491	0.0000	0.0000
*** LIQUID PHASE ***					
RATE LB /HR	6918.8027	16215.1074	0.0000	249.9999	242.5315
ACT.RATE BBL/DAY	758.47	1446.55	0.00	17.34	16.82
STD. LV RATE BBL/HR	21.98	59.02	0.00	0.72	0.69
CP, BTU /LB F	0.6924	0.4958	0.0000	0.9929	1.1699
MOLECULAR WEIGHT....	199.5084	136.7552	0.0000	18.0150	18.0087
ACT.DENS LB /FT3	39.9930	47.9157	0.0000	61.6306	61.6314
STD. API GRAVITY....	25.7300	42.6594	0.0000	10.0635	10.2821
*** DRY BASIS ***					
RATE LB /HR	6918.5146	16210.8398	0.0000	0.0000	0.3195
MOLECULAR WEIGHT....	199.5922	136.9928	0.0000	0.0000	14.2259
UOP K	11.1656	11.5915	0.0000	0.0000	13.0285
FLASH POINT, DEG F	19.0925	-66.1856	0.0000	0.0000	1.9601
CRIT. TEMP, F	722.4723	584.6508	0.0000	0.0000	6.5419
CRIT. PRES, PSIA	271.2542	367.0006	0.0000	0.0000	1028.8279
*** VAPOR PHASE ***					
RATE LB /HR	0.0000	0.0000	2293.0903	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	1.20	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.30	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	2.5153	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	2.9014	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.5315	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	1.0492	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***					
RATE LB /HR	6918.5146	16210.8398	0.0000	0.0000	0.3195
ACT.RATE BBL/DAY	758.43	1446.26	0.00	0.00	0.04
CP, BTU /LB F	0.6923	0.4936	0.0000	0.0000	0.7504
MOLECULAR WEIGHT....	199.5922	136.9925	0.0000	0.0000	14.2259
ACT.DENS LB /FT3	[illegible]	47.9129	0.0000	0.0000	31.7042
STD. API GRAVITY....	[illegible]	[illegible]	0.0000	0.0000	175.9743

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	25	26	27	28	30	31
STREAM NAME..........	HDC NAPHTHA	HDC JP-8	FRAC CVHD	HDC HC VAP	FRAC STR STM	JP-8 STR STM
STREAM PHASE.........	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F	110.0000	354.9146	291.1038	110.0000	318.6042	311.6954
PRESSURE, PSIA	150.0000	55.0000	60.0000	50.0000	67.5000	55.0000
RATE LB MOLS/HR	58.2450	36.3234	222.2137	16.2545	48.2931	15.2651
RATE LB /HR	5649.6533	5929.9980	15413.7305	295.7139	869.9995	274.9999
ENTHALPY MM BTU /HR	0.0985	0.8636	4.8493	0.0359	1.0352	0.3272
ENTHALPY BTU /LB	17.4365	145.6345	314.6068	121.5405	1189.8687	1189.8687
MOLECULAR WEIGHT....	96.9982	163.2557	69.3644	18.1927	18.0150	18.0150
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	0.0000	15413.7305	295.7139	869.9995	274.9999
ACT.RATE FT3/SEC	0.00	0.00	7.91	0.55	1.60	0.62
STD.RATE MM FT3/DAY	0.00	0.00	2.02	0.15	0.44	0.14
CP, BTU /LB F	0.0000	0.0000	0.5339	0.6564	0.5462	0.5305
MOLECULAR WEIGHT....	0.0000	0.0000	69.3644	18.1927	18.0150	18.0150
ACT.DENS LB /FT3	0.0000	0.0000	0.5412	0.1494	0.1513	0.1236
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9545	0.9960	0.9623	0.9685
*** LIQUID PHASE ***						
RATE LB /HR	5649.6533	5929.9980	0.0000	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY	571.89	606.46	0.00	0.00	0.00	0.00
STD. LV RATE BBL/HR	23.07	21.40	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5374	0.6237	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	96.9982	163.2557	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	42.2281	41.7968	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	70.5799	47.0949	0.0000	0.0000	0.0000	0.0000
*** DRY BASIS ***						
RATE LB /HR	5647.9697	5916.9121	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	97.1252	166.2198	0.0000	0.0000	0.0000	0.0000
UOP K	12.4155	12.0317	0.0000	0.0000	0.0000	0.0000
FLASH POINT, DEG F	-40.3827	96.6095	0.0000	0.0000	0.0000	0.0000
CRIT. TEMP, F	490.8882	718.5243	0.0000	0.0000	0.0000	0.0000
CRIT. PRES, PSIA	437.1967	322.2452	0.0000	0.0000	0.0000	0.0000
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	0.0000	14294.7852	288.2875	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	5.62	0.54	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.06	0.01	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.5428	0.6616	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	89.2869	18.1973	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.7068	0.1494	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9403	0.9960	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0097	0.0122	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	5647.9697	5916.9121	0.0000	0.0000	0.0000	0.0000
ACT.RATE BBL/DAY	571.77	605.45	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5373	[illegible]	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	97.1252	166.2198	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	42.2241	41.7741	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	[illegible]	[illegible]	[illegible]	0.0000	0.0000	0.0000

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	32	33
STREAM NAME.........	SOUR H2O	REFLUX
STREAM PHASE........	LIQUID	LIQUID
TEMPERATURE, DEG F	110.0000	110.0000
PRESSURE, PSIA	50.0000	50.0000
RATE LB MOLS/HR	[illegible]	86.3171
RATE LB /HR	1107.3677	8369.8887
ENTHALPY MM BTU /HR	0.0863	0.1463
ENTHALPY BTU /LB	77.9702	17.4835
MOLECULAR WEIGHT....	18.0150	96.9667
*** VAPOR PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	1107.3677	8369.8887
ACT.RATE BBL/DAY	76.53	848.50
STD. LV RATE BBL/HR	3.17	34.17
CP, BTU /LB F	0.9978	0.5373
MOLECULAR WEIGHT....	18.0150	96.9667
ACT.DENS LB /FT3	61.8534	42.1556
STD. API GRAVITY....	10.0635	70.5916
*** DRY BASIS ***		
RATE LB /HR	0.0000	8367.3945
MOLECULAR WEIGHT....	0.0000	97.0937
UOP K	0.0000	12.4155
FLASH POINT, DEG F	0.0000	-40.9109
CRIT. TEMP, F	0.0000	490.7809
CRIT. PRES, PSIA	0.0000	437.2733
*** VAPOR PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	0.0000	8367.3945
ACT.RATE BBL/DAY	0.00	[illegible]
CP, BTU /LB F	0.0000	0.5372
MOLECULAR WEIGHT....	0.0000	[illegible]
ACT.DENS LB /FT3	0.0000	42.1515
STD. API GRAVITY....	[illegible]	[illegible]

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	40	41	42	43	44
STREAM NAME..........	NAPHTHA FEED	STAB OFFGAS	REFLUX	NAPHTHA PROD	REBOILER
STREAM PHASE.........	LIQUID	VAPOR	LIQUID	LIQUID	LIQUID
TEMPERATURE, DEG F	97.8588	109.2971	109.2971	372.6451	316.1483
PRESSURE, PSIA	144.0000	125.0000	125.0000	142.0000	141.5000
RATE LB MOLS/HR	150.8883	7.5061	18.7648	143.2816	308.7294
RATE LB /HR	14776.7715	319.9324	1045.5327	14455.0059	29079.9414
ENTHALPY MM BTU /HR	0.0721	0.0679	0.0451	2.4428	3.9098
ENTHALPY BTU /LB	4.8822	212.3046	43.1618	168.9918	134.4507
MOLECULAR WEIGHT....	97.9319	42.6228	55.7178	100.8853	94.1923
*** VAPOR PHASE ***					
RATE LB /HR	0.0000	319.9324	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.09	0.00	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.07	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.4313	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	42.6223	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.9854	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.8856	0.0000	0.0000	0.0000
*** LIQUID PHASE ***					
RATE LB /HR	14776.7715	0.0000	1045.5327	14455.0059	29079.9414
ACT.RATE BBL/DAY	1422.75	0.00	133.44	1793.74	3478.13
STD. LV RATE BBL/HR	57.94	0.00	5.22	56.26	115.13
CP, BTU /LB F	0.5085	0.0000	0.7083	0.7027	0.6870
MOLECULAR WEIGHT....	97.9319	0.0000	55.7178	100.8853	94.1923
ACT.DENS LB /FT3	44.3959	0.0000	33.4915	34.4469	35.7388
STD. API GRAVITY....	62.5768	0.0000	115.3759	61.1234	64.4487
*** DRY BASIS ***					
RATE LB /HR	14773.3887	0.0000	1044.9988	14454.7793	29075.9336
MOLECULAR WEIGHT....	98.0315	0.0000	55.7775	100.8926	94.2474
UOP K	12.0088	0.0000	13.6353	11.9656	12.0326
FLASH POINT, DEG F	-22.1483	0.0000	-115.0079	-7.3064	-38.9144
CRIT. TEMP, F	515.6772	0.0000	284.2679	530.8590	500.6957
CRIT. PRES, PSIA	465.1137	0.0000	587.5289	448.3092	468.9988
*** VAPOR PHASE ***					
RATE LB /HR	0.0000	318.5854	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.09	0.00	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.4818	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	42.8699	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.9923	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.8845	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0091	0.0000	0.0000	0.0000
*** LIQUID PHASE ***					
RATE LB /HR	14773.3887	0.0000	1044.9988	14454.7793	29075.9336
ACT.RATE BBL/DAY	1422.52	0.00	133.41	1793.72	3477.83
CP, BTU /LB F	0.5094	0.0000	0.7091	0.7027	0.6870
MOLECULAR WEIGHT....	98.0315	0.0000	55.7775	100.8926	94.2474
ACT.DENS LB /FT3	[illegible]	0.0000	33.4936	34.4467	35.7369
STD. API GRAVITY....	[illegible]	[illegible]	[illegible]	61.1242	64.4562
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

APPENDIX E

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 4
PROFITABLE JP-8 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

TABLE OF CONTENTS

1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to produce JP-8 type aviation turbine fuel and chemical byproducts to maximize profit from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- Tar Oil byproduct stream (47620 #/hr, 3182 BPSD) is charged to the hydrotreater (Area 100).
- The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 550^{o}F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (4100 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.
- The hydrotreater produces 6 streams:
  - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.
  - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
  - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to storage and gasoline blending.
  - JP-8 turbine fuel which is combined with JP-4 produced in the hydrocracker (area 200) and sent to storage.
  - 550^{o}F+ unconverted bottoms product which is sent to the hydrocracker (area 200).
  - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

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- Approximately 950 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.

. The 550^{o}F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-8 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.

. The hydrocracker produces 4 streams in addition to JP-8

- High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H_2 and CH_4.

- Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

- Stabilized naphtha which is sent to storage and gasoline blending.

- A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.

. Hydrogen make-up for the Hydrotreater, the Hydrocracker and the Naphtha Hydrotreater is supplied from a PSA Hydrogen Unit (Area 300). High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft^3. This H_2, CO & CH_4 rich gas is recompressed into the methanation unit of the SNG plant.

. The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).

. The distillation removes the material boiling below 160^{o}F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.

. The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.

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1.1 <u>Overall Process Description</u> - cont'd

- The naphtha hydrotreater produces 4 streams:

  - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

  - Naphtha which is stabilized to control vapor pressure, and the sent to the aromatics recovery unit (Area 700).

  - A low pressure off gas which is sent to the Stretford unit in the SNG plant.

  - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

- The hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.

- Five streams are produced in the ARU plant.

  - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.

  - A small process water stream which is sent to the waste treatment plant in the SNG Plant.

  - Three product streams Benzene, Toluene & Xylene which are sent to storage.

- The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).

1.1 Overall Process Description - cont'd

- Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.

- The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.

- The acid tar produced is water washed and mixed with light oil and sent to fuel.

- The remaining cresol/xylenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).

- Streams produced in the phenol extraction unit are:

  - Phenol product sent to storage

  - Crude Cresylic Acid sent to distillation (Area 900) or storage.

  - Wash Water sent to Water Treatment in the SNG Plant.

  - Waste Water sent to the Phenosolvan unit in the SNG Plant.

  - Neutral Oil sent to storage and fuel for the SNG Plant boilers.

- The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.

- Streams produced in the crude cresylic acid distallation unit are:

  - o-Cresol product which is sent to storage.

  - m,p-Cresol product which is sent to storage.

  - Xylenol product which is sent to storage.

  - A heavy distillate which is combined with neutral oil in Area 800.

1.1 Overall Process Description - cont'd

- A crude phenol stream which is recycled to the Area 800.

- A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas, neutral oil and 160°F-distillate produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

936 BPSD of Crude Phenol
725 BPSD of Crude Naphtha
3182 BPSD of Tar Oil
4361 BPSD of #6 Fuel Oil
10.76 MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

2552 BPSD of JP-8 turbine fuel
1276 BPSD of 300°F - Naphtha for gasoline blending
317 BPSD of Phenol
56 BPSD of o-Cresol
131 BPSD of m,p-Cresol
75 BPSD of Xylenols
312 BPSD of Neutral Oil for Fuel
202 BPSD of 160°F - Distillate for Fuel
46 BPSD of Gasoline Blending Stock
315 BPSD of Benzene
112 BPSD of Toluene
15 BPSD of Xylene
7.02 MMSCFD equivalent SNG product credit due to HDT, HDC & PSA purge gas reinjection into SNG plant.

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1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	4361 BPSD
SNG Equivalent	
of Syn Gas & Purge Gas	3.73 MM SCFD
Power	6160 kW
Cooling Water	6050 GPM (30^{o}F rise)
Process Water	24 GPM

In addition the process utilizes steam as summarized below which was debited against boiler requirements.

HP Steam Import	58,200 #/H
MP Steam Import	8,900 #/H
LP Steam Export	6,900 #/H
Condensate Return	67,100 #/H

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Figure 1 Case 4:Profitable JP-8

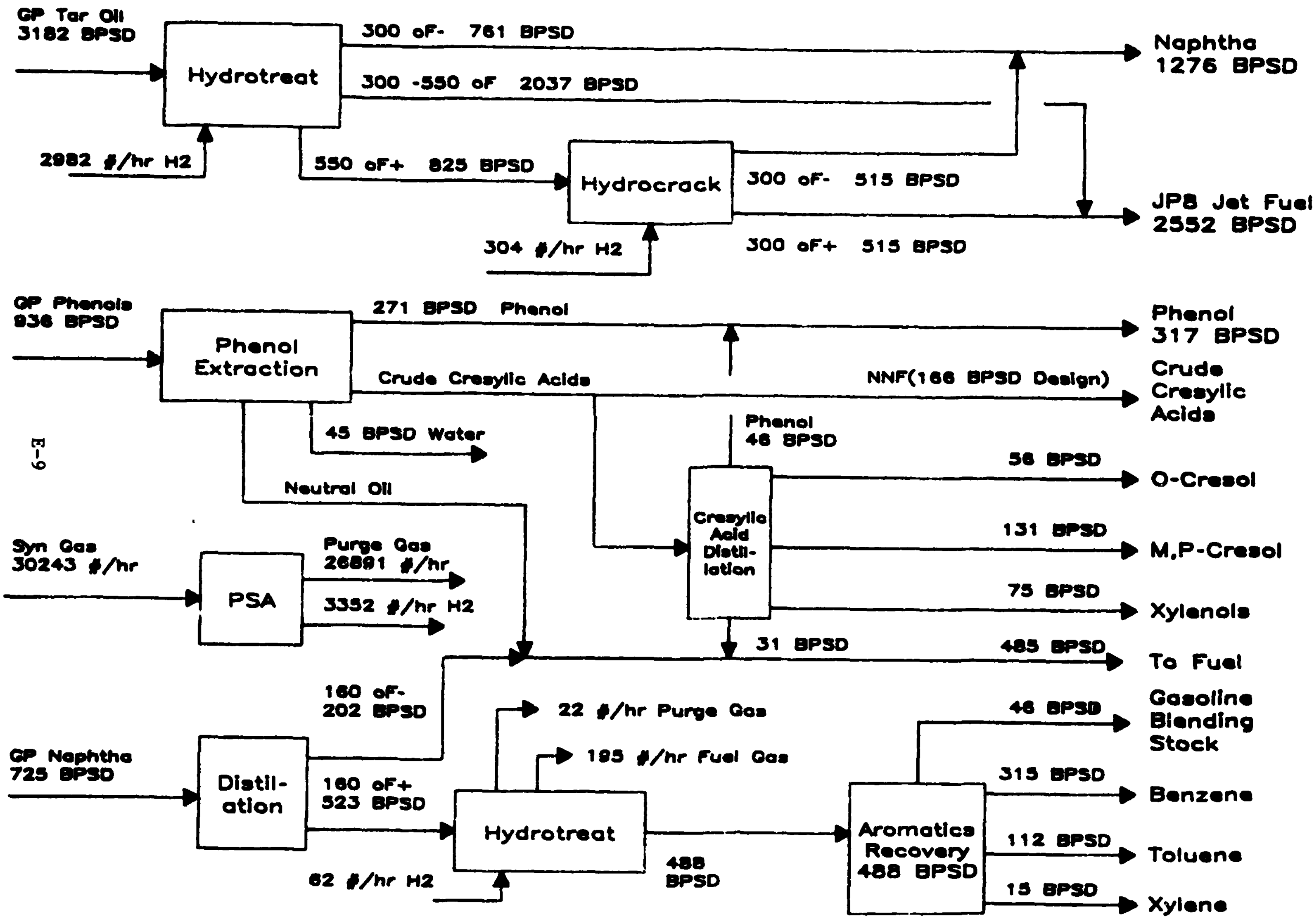


Table 1.1 Great Plains Case 4: Profitable JP8 Production

Tar Oil Feed=====>	47620	#/hr	3182	BPSD
Phenol Feed======>	14490	#/hr	936	BPSD
Crd Naphtha Feed=>	8738	#/hr	725	BPSD
Naphtha Product==>	13588	#/hr	1276	BPSD
JP8 Product======>	30631	#/hr	2552	BPSD
Phenol Product===>	4925	#/hr	317	BPSD
o-Cresol Prod====>	845	#/hr	56	BPSD
m,p-Cresol Prod==>	1974	#/hr	131	BPSD
Xylenols Prod====>	1070	#/hr	75	BPSD
Gasoline Stock===>	481	#/hr	46	BPSD
Benzene Prod=====>	4060	#/hr	315	BPSD
Toluene Prod=====>	1425	#/hr	112	BPSD
Xylene Prod======>	188	#/hr	15	BPSD
SNG Product Loss=>	6411	#/hr	3.7	MMSCFD
Fuel Oil Makeup==>	60370	#/hr	4361	BPSD

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feeds					
H2	6.26		2982	1479.2	
Tar Oil	100.00	1.0268	47620		3182
Total	106.26		50602		
Products					
Purge Gas	0.10		48	14.9	
Fuel Gas	1.87		891	43.9	
Naphtha	17.70	0.7600	8430		761
JP-8	51.87	0.8320	24701		2037
550 oF+	24.49	0.9700	11663		825
H2O in SW	8.73		4159	231.1	
H2S in SW	0.43		205	6.0	
NH3 in SW	1.06		505	29.7	
Total	106.26		50602		3623

Fixed Bed Hydrocracker

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feeds					
H2	2.61		304	150.8	
500oF+	100.00	0.9700	11663		825
Total	102.61		11967		
Products					
Purge Gas	1.48		173	58.6	
Fuel Gas	4.27		498	27.5	
Naphtha	46.00	0.7148	5365		515
JP-8	50.84	0.7900	5930		515
H2S in SW	0.003		0.4	0.01	
NH3 in SW	0.003		0.4	0.02	
Total	102.60		11967		1030

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
HDT Nap	61.11	0.7600	8430		761
HCR Nap	38.89	0.7148	5365		515
Stab Nap	98.50	0.7306	13588		1276
Fuel Gas	1.50		207	4.8	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1931.1	569.0	45.2	495.6	9.6	5.8	3056.2
Prod. H2	1660.7	0.2	0.0	0.0	0.0	0.0	1660.9
Purge Gas	270.4	568.8	45.2	495.6	9.6	5.8	1395.3
#/hr							
Feed Gas	3893	15935	1990	7950	289	186	30243
Prod. H2	3348	4	0	0	0	0	3352
Purge Gas	545	15931	1990	7950	289	186	26891

Crude Naphtha Distillation	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

Aromatics Recovery

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	6154	488
Products				
Raffinate	7.82	0.7175	481	46
Benzene	65.97	0.8844	4060	315
Toluene	23.16	0.8718	1425	112
Xylene	3.05	0.8729	188	15
Total Products	100		6154	488

Phenol Extraction

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Cr. Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	891	43.9	16.0
HCR FG Produced	498	27.5	9.0
Stabilizer FG	207	4.8	3.7
Naphtha Hdtr FG	195	10.8	3.5
Total Fuel Gas	1596	76.3	28.7

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	545	270.4	324	33.2
CO	15931	568.8	321	69.2
CO2	1990	45.2	0	0.0
C1	7950	495.6	1010	189.7
C2	289	9.6	1769	6.5
N2+Ar	186	5.8	0	0.0
Total	26891	1395.3	565	298.6

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Tar Oil	-47620	17000	-809.5			-3182
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
Fuel Gas	1596	18000	28.7	0.7	994	
160 oF- distillate	2164	17400	37.7			202
Neutral Oil	4936	15000	74.0			312
Import Steam	-60200	1000	-60.2			
Fuel Oil to Boiler	60020	18000	1080.4			4336
Total	-62332		0.0	0.7		7
Fuel Oil to Process Heaters	350	18000	6.3			25

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	10.76	73350
SNG Credit for PSA Purge gas	6.74	33488
SNG Credit for Hdtrs purge gas	0.29	1929
Total SNG Production Loss	3.73	

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

For a description of the Hydrotreater process see Case 3 Section 2.1.

2.2 Hydrocracker (Area 200)

For a description of the Hydrocracker process see Case 3 Section 2.2.

2.3 PSA Hydrogen Unit & Recompression (Area 300)

2.3.1 Hydrogen for both the hydrotreater and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 ^{o}F
Composition	mol%

H_2	63.19
CO	18.61
CO_2	1.48
CH_4	16.21
C_2H_6	0.31
COS, H_2S, CS_2	< 0.01
N_2 + Ar	0.19
H_2O	< 0.01

The PSA unit selectively absorbs all components expect H_2 and produces a 99.99% vol. purity stream at about 345 psig and 80^{o}F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100^{o}F
Composition	Mole %
H_2	19.32
CO	40.76
CO_2	3.24
CH_4	35.51
C_2H_6	0.69
N_2+Ar	0.41
Others	0.06

At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

2.3 PSA Hydrogen Unit & Recompression (Area 300) - cont'd

2.3.1 Cont'd

The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-40301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

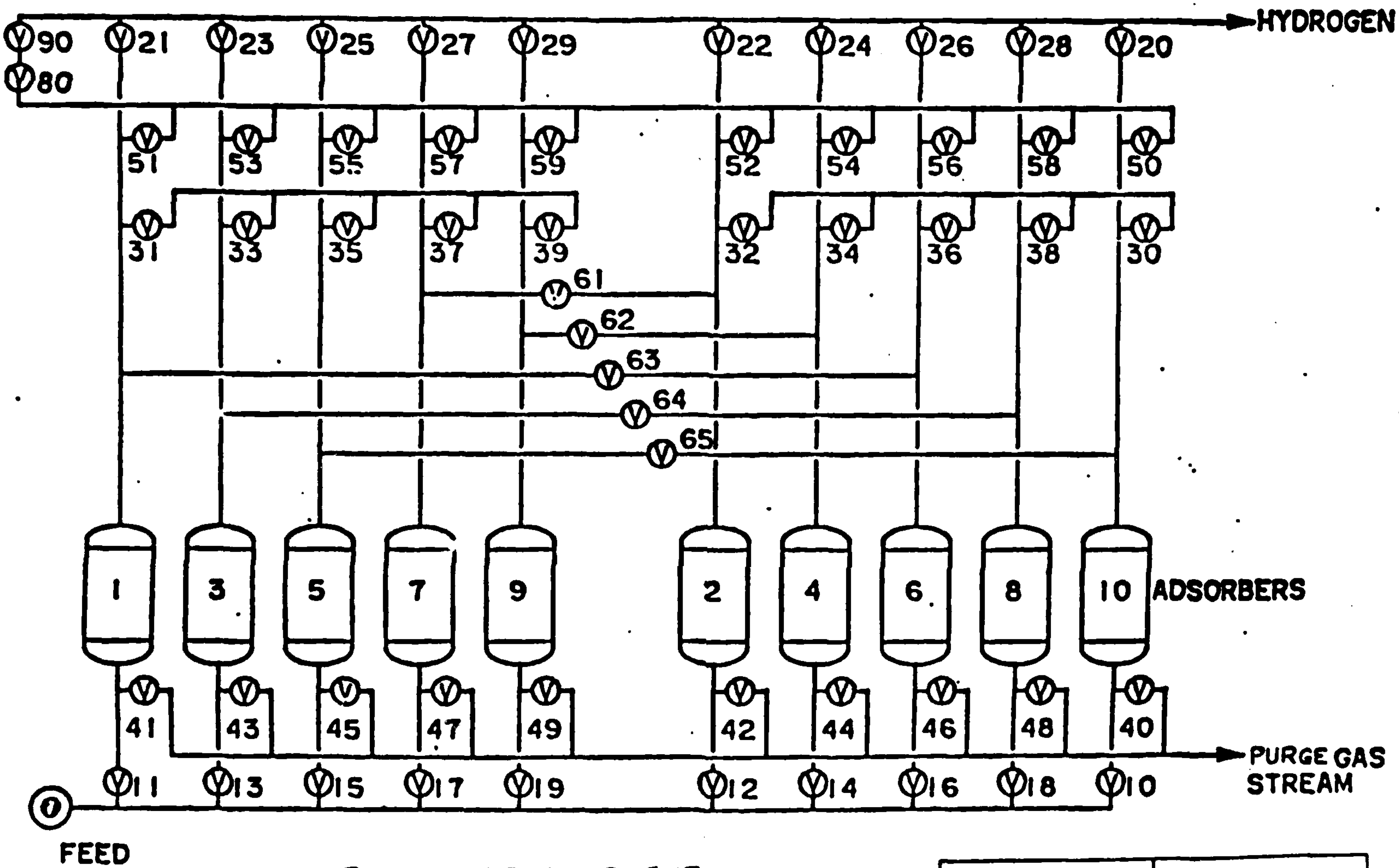
2.4 Phenol Stream (Areas 800 and 900)

For a description of the Phenol Extraction (Area 800) and Cresylic Acid Distillation (Area 900) Units see Case 7 Section 2.1.

2.5 Naphtha Stream (Area 600 and 700)

For a description of the Naphtha Distillation and Hydrotreating Unit (Area 600) and the Aromatics Recovery Unit (Area 700) see Case 7 Section 2.2.

SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 10

REV	ISSUE DATE	DESCRIPTION	PROC ENGR	PROJ ENGR	APPR	APPR
△						
△	1/31/80	FOR SUBTASK 1.2	ML	E.S.		

LUMMUS — THE LUMMUS COMPANY

TITLE PSA HYDROGEN UNIT
AREA 300

CLIENT AMOCO/DOE

PROJ NO 5571

CASE 4

DWG NO - 5571-40301

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 4 - PROFITABLE JP-8

AREA 100 - HYDROTREATER

TAG. NO.	DESCRIPTION
See Case 3 Area 100	

AREA 200 - HYDROCRACKER

See Case 3 Area 200

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

AREA 400 - STORAGE AREA

FB-401	Jet Fuel Storage Tank
FB-402	Naphtha Storage Tank
FB-403	Fuel Oil Storage Tank
FB-404	Blending Stock
FB-405	Benzene Storage
FB-406	Toluene Storage
FB-407	Xylene Storage
FB-409	Gasoline Storage
FB-410	Neutral Oil Storage
FB-411	Phenol Product Storage
FB-412	Crude Cresylic Acid Storage
FB-413	O-Cresol Storage
FB-414	M, P Cresol Stoage
FB-415	Xylenol Storage

3.0 CAPITAL COSTS

3.1 Equipment List - cont'd

CASE 4 - PROFITABLE JP-8

TAG NO.		DESCRIPTION
AREA 400		STORAGE AREA
GA-401A/S		Tar/Tar Oil Feed Pump
GA-402A/S		Crude Phenol Feed Pump
GA-403A/S		Fuel Oil Transfer Pump
GA-404A/S		Naphtha Transfer Pump
GA-405A/S		Crude Naphtha Transfer Pump
GA-406A/S		Blending Stock Pump
GA-407A/S		Benzene Transfer Pump
GA-408A/S		Toluene Transfer Pump
GA-411A/S		Gasoline Transfer Pump
GA-412A/S		Neutral Oil Transfer Pump
GA-413A/S		Crude Cresylic Acid Transfer Pump
GA-414A/S		O-Cresol Transfer Pump
GA-415A/S		M, P. Cresol Transfer Pump
GA-416A/S		Xylenol Transfer Pump
PA-401		Gasoline Blending Package
AREA 500	-	CATALYST HANDLING
See Case 3 Area 500		
AREA 600	-	NAPHTHA DISTILL. AND HDT
See Case 7 Area 600		
AREA 700	-	AROMATICS RECOVERY
See Case 7 Area 700		
AREA 800	-	PHENOL EXTRACTION
See Case 7 Area 800		
AREA 900	-	CRESYLIC ACID DISTILLATION
See Case 7 Area 900		

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate for this case is a factored type estimate using the T.I.C. values developed for the various cases referenced in this project.

The total investment costs are scaled to the capacity requirement of ths case use a o.6 exponent.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs for Areas 100 thru 700 and 30% for Areas 800 & 900.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of $)

	Case 4
Area 100 Hydrotreater	$20,702
Area 200 Hydrocracker	10,012
Area 300 PSA & Recompression	8,400
Area 400 OSBL	9,421
Area 500 Catalyst Handling	1,285
Area 600 Naph. Dist & HDT	4,615
Area 700 ARU	7,887
Area 800 Phenol Ext.	12,276
Area 900 Cresylic Acid Dist.	4,832
Subtotal	$79,430
Area 700 ARU Solvent Inventory	100
Total	$79,530

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands $

This unit has the same capacity as the 100 Area of Case 3. Therefore, T.I.C. = $20,702.

Area 200

This unit has a 100% capacity of the 200 Area of Case 3. Therefore, T.I.C. = $10,012.

W16245/bf

3.2.3 Estimate Breakdown - Cont'd

Area 300

This unit has a 102% capacity of the 300 Area of Case 3. Therefore T.I.C. = $(1.02)^{0.6}$ (8300) = \$8400.

Area 400

Tar Oil Stream Storage 100% Case 1
TIC = \$5,110

Phenol Stream Storage = 100% Case 7
TIC = \$3,016

Naphtha Stream Storage 100% Case 7
TIC = \$3058

	Subtotal	\$11,184
Less Duplicate Pipe & Rack		-1763
	Total =	\$9,421

Area 500

The capacity of this unit is identical to the 500 Area of Case 3. Therefore T.I.C. = \$1,285

Area 600

This unit has a capacity of 100% of the 600 Area of Case 7. Therefore T.I.C. = \$4,615

Area 700

This unit has a capacity of 100% of the 700 Area of Case 7. Therefore T.I.C. = \$7,887

Area 800

This unit has a capacity of 100% of the 800 Area of Case 7. Therefore T.I.C. = \$12,276

Area 900

This unit has a capacity of 100% of the 900 Area of Case 7. Therefore = T.I.C. = \$4,832

W16245/bf

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require men/shift to operate the plant broken down as follows:

Foreman	2
Control Room	2
HDT Operator	2
HCR Operator	2
PSA & relief man	1
Naphtha Distil. & HDT	2
ARU	2
Phenol Ext.	1
Cresylic Acid Dist.	1
	15 Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 7 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	15 positions x 4 people/position -	60
Supervisor & Admin.		6
QC Technician		2
Maintenance		7
Other (Stores or Janitorial)		1
Total		76

4.2 Utilities

The following utilities have been estimated:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	4361 BPSD	$16/Bbl [a]	69776
SNG equivalent of Syn Gas & Purge Gas	3.73 MMSCFD	$3.80/MM Btu [b]	13892
Cooling Water	6050 GPM	$0.155/MGal [c]	1350
Power	6160 kW	$0.04/kWH [c]	5914
Process Water	24 GPM	$0.45/MGal [c]	15

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

4.2 Utilities - cont'd

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

Catalyst & Chem.	Use	Cost	$/SD
Nap. HDT Cal	0.021 #/Bbl	$3.00/#	33
HDT Cat.	0.30 #/Bbl	$3.00/#	2864
HCR Cat.	0.0095 #/Bbl	$6.00/#	47
Inhibitors	50 PPM	$10/Gal	52
ARU Solvent	24 #/D	$2.10/#	50
H_2SO_4	7100 #/D	$0.04/#	285
			$3331

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units (excluding the ARU solvent inventory). On this basis the maintenance supplies would be

$$0.00005 \times 79,430,000 = \$3971/SD$$

W16245/bf

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-8 and by-product chemicals are proposed to be located to the east of the Phenosolvan Unit and Water Treatment Area of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 340' x 575' will be surrounded by an access road and will be divided by three central east-west roads. Areas 100 & 500 will be located to the north and Areas 200 & 300 south of Area 100, Areas 800 & 900 next and then Areas 600 & 700.

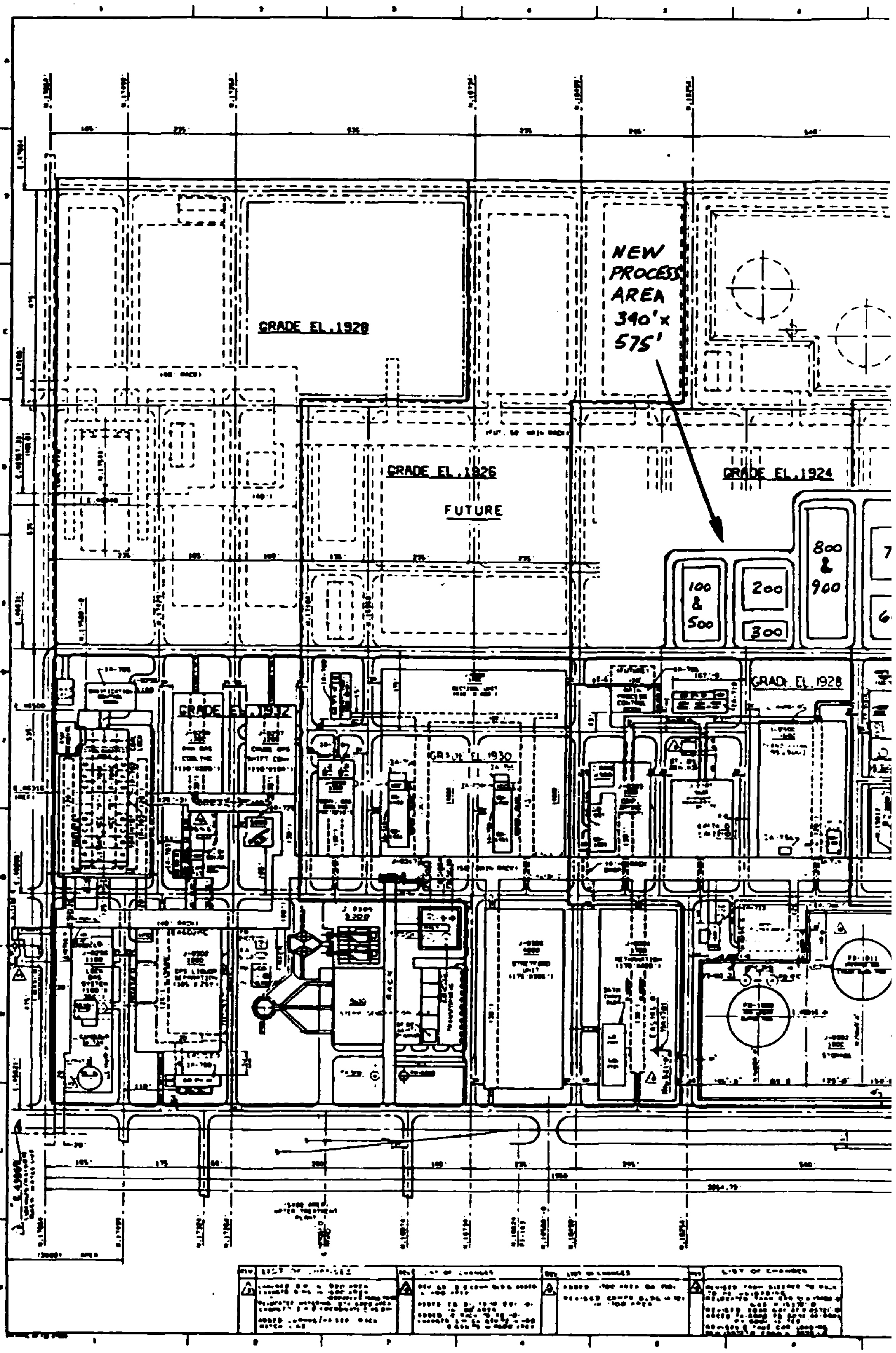
A diked storage tank area approx. 375' x 425' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

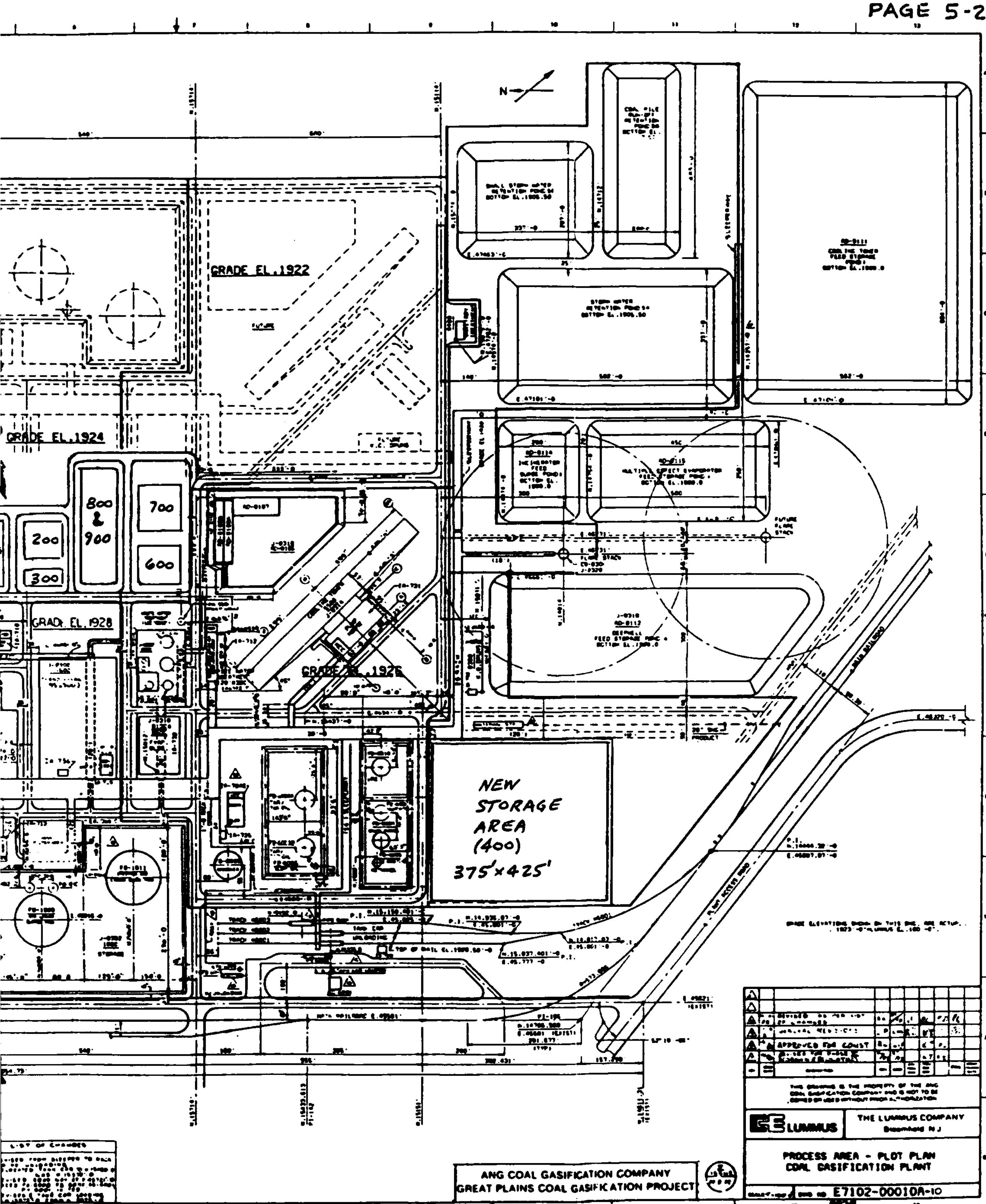
5.2 Unit Tie-Ins

Approximately 2500 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

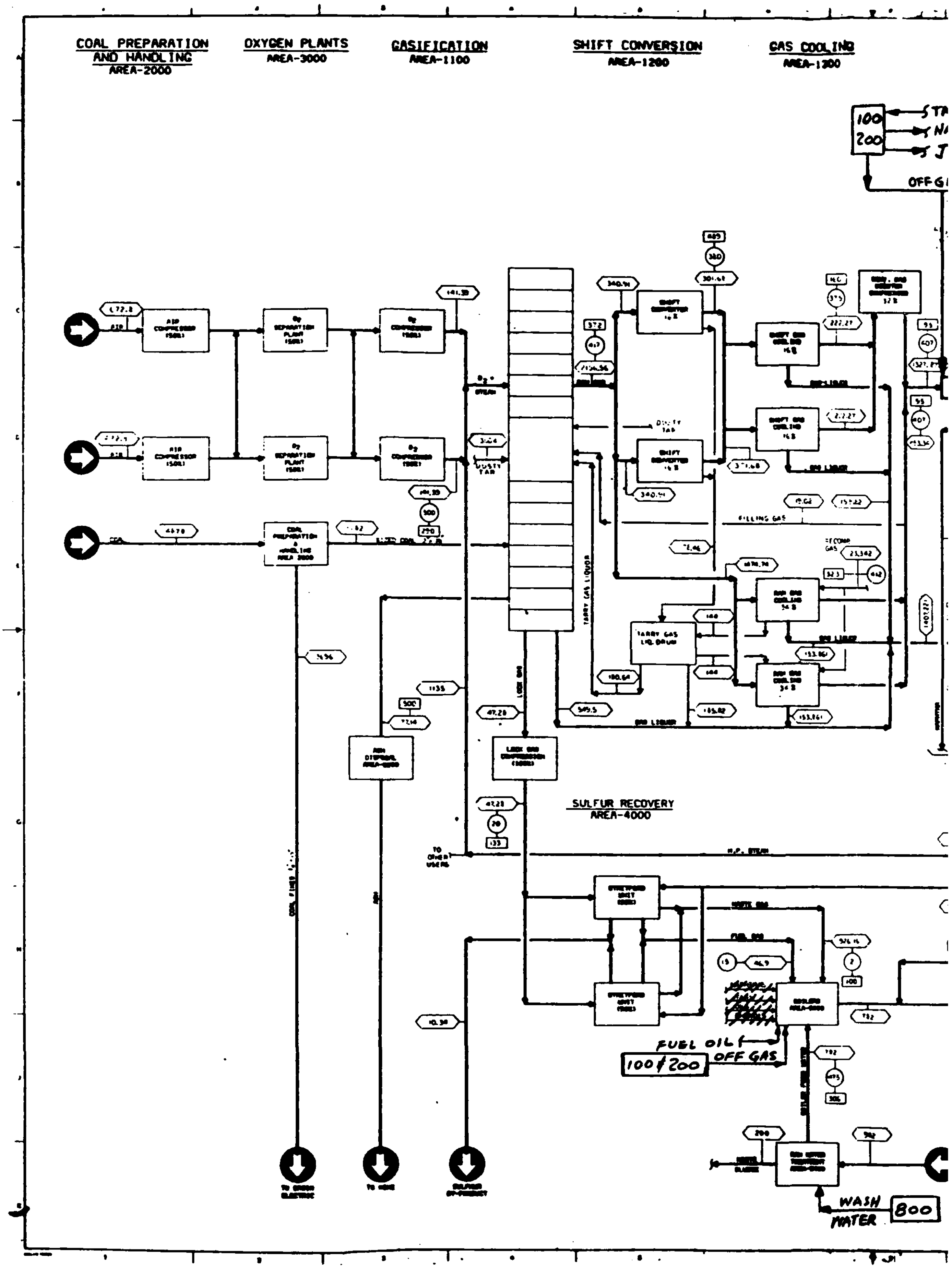
New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

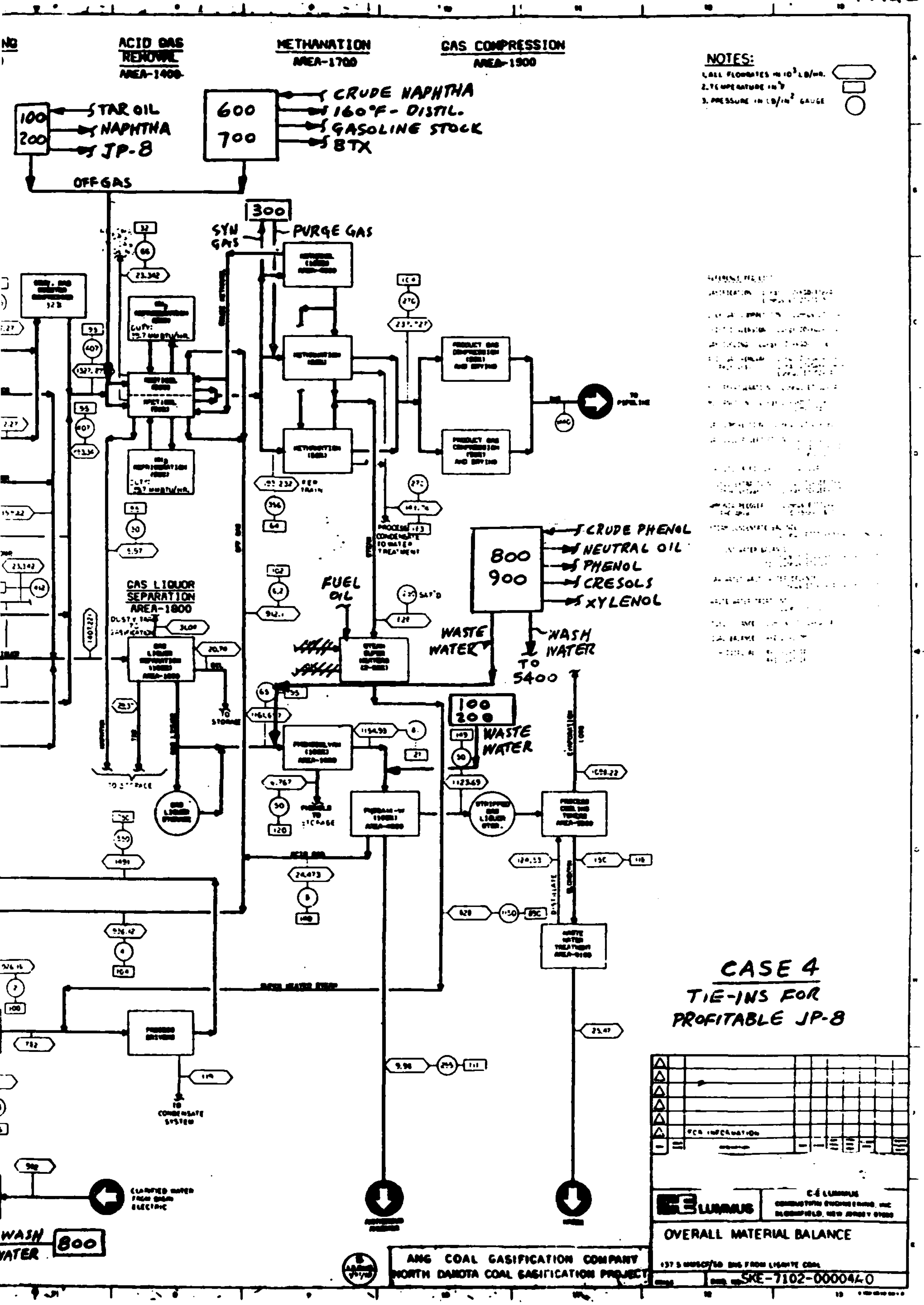
A summary of the lines has not been prepared for this case but will be similar to a combination of Cases 3 and 7 with the utility lines of like services being combined.


NEW PROCESS AREA 340'x 575'
GRADE EL.1928
GRADE EL.1926
FUTURE
GRADE EL.1924
100 & 500
200
300
800 & 900
GRAD. EL.1928
GRADE EL.1932
GRADE EL.1930


N
GRADE EL.1922
GRADE EL.1924
GRADE EL.1928
GRADE EL.1926
800 & 900
700
200
600
300
NEW STORAGE AREA (400) 375'×425'
ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT
LUMMUS
THE LUMMUS COMPANY
PROCESS AREA - PLOT PLAN
COAL GASIFICATION PLANT
E7102-00010A-10

CASE 4


COAL PREPARATION AND HANDLING
AREA-2000
OXYGEN PLANTS
AREA-3000
GASIFICATION
AREA-1100
SHIFT CONVERSION
AREA-1200
GAS COOLING
AREA-1300
100
200
OFF GAS
SULFUR RECOVERY
AREA-4000
FUEL OIL
100+200
OFF GAS
WASH WATER
800


ACID GAS REMOVAL AREA-1400
METHANATION AREA-1700
GAS COMPRESSION AREA-1900
NOTES:
100
200
TAR OIL
NAPHTHA
JP-8
600
700
CRUDE NAPHTHA
160°F - DISTIL.
GASOLINE STOCK
BTX
OFF GAS
300
SYN GAS
PURGE GAS
TO PIPELINE
GAS LIQUOR SEPARATION AREA-1800
FUEL OIL
WASTE WATER
800
900
CRUDE PHENOL
NEUTRAL OIL
PHENOL
CRESOLS
XYLENOL
WASH WATER
TO 5400
100
200
WASTE WATER
TO STORAGE
TO CONDENSATE SYSTEM
CLARIFIED WATER FROM BASIN ELECTRIC
WASH WATER
800
CASE 4
TIE-INS FOR
PROFITABLE JP-8
C-E LUMMUS
OVERALL MATERIAL BALANCE
ANG COAL GASIFICATION COMPANY
NORTH DAKOTA COAL GASIFICATION PROJECT
SKE-7102-0000440

APPENDIX F

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 5
PROFITABLE JP-8X PRODUCTION
SUBTASKS 1.2

LCI PROJECT 5571
DATE - JAN. 30, 1988

W16248/bf

1.0 CASE DESCRIPTION

The design basis for the Maximum JP-8X case was received from Amoco and is presented on the following pages.

For various reasons, the most significant being the time restraint, this case was not developed into a conceptual process design and estimated by Lummus.

Also, of significance in the decision was the importance of developing Case 7 - Maximum Profit and that the Amoco LP did not choose vacuum distillation for the Profitable JP-8X Case 6.

FIGURE 1
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 5: MAXIMUM JP-8X PRODUCTION

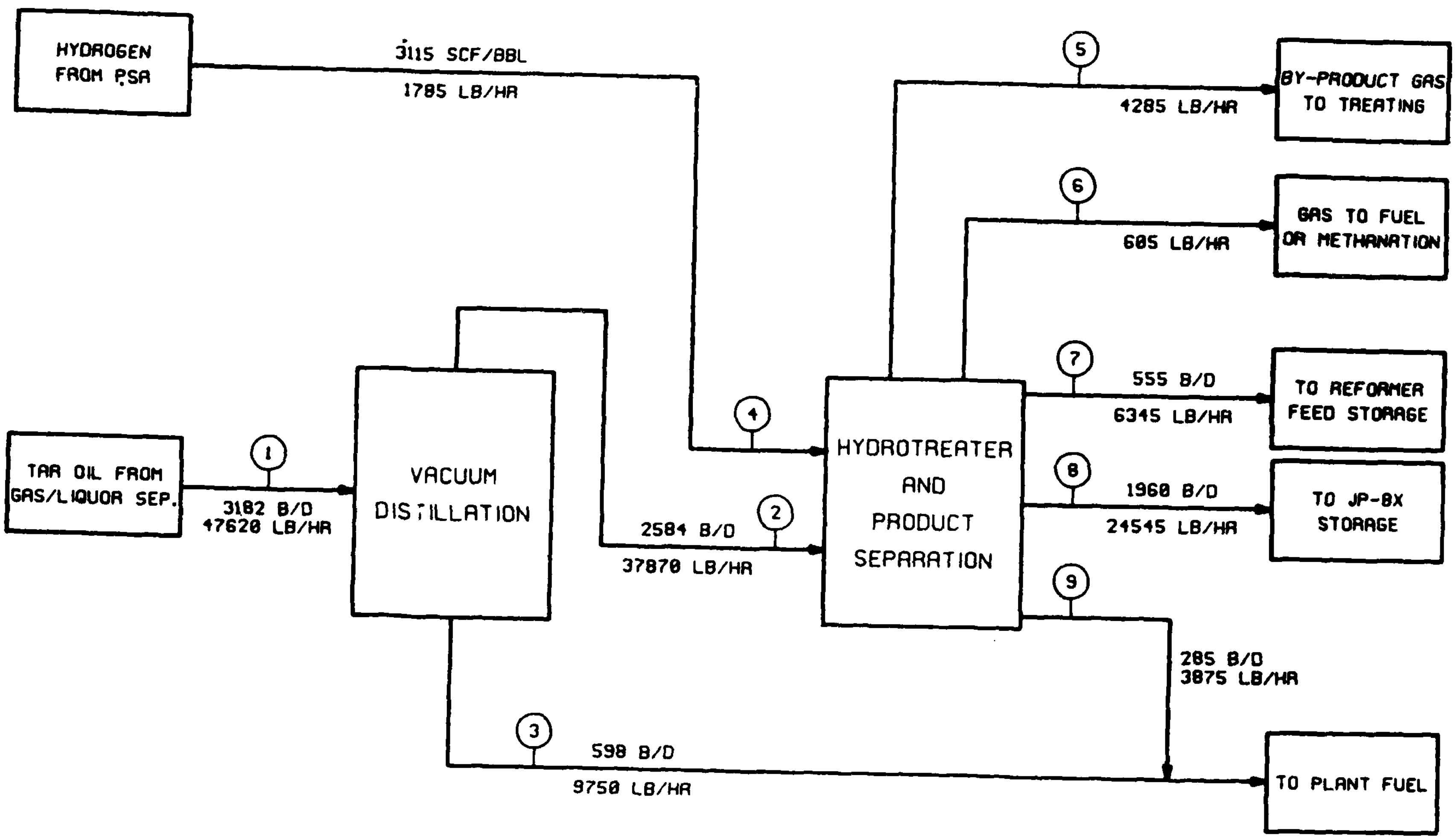


TABLE I

TAR OIL PROPERTIES
GREAT PLAINS CASE 5:
MAXIMUM JP-8X PRODUCTION

Flow Rate, B/D	3,182±500
API	6.6
Elemental Analysis, Wt%:	
C	82.1
H	8.6
N	0.9
S	0.4
O (Difference)	8.0
Solids, Wt%	0.8±0.6
Ash, Wt%	0.1±0.08
Ramscarbon Residue, Wt%	2.0
Water, Wt%	2.0±1.0
Simulated Distillation*	
Temperature @ IBP, °F	225
5%	320
10%	352
20%	405
30%	454
50%	542
70%	656
80%	734
90%	845
95%	918
FBP	1018

*D-2887 with aromatic standard

LPS/lkg/87863
9/9/87

TABLE II

STREAM FLOWS AND COMPOSITION
GREAT PLAINS CASE 5:
MAXIMUM JP-8X PRODUCTION

Stream No.	1	2	3	4	5	6	7	8	9
Description	Tar Oil	750-°F Tar Oil	750+°F Tar Oil	Hydrogen	By-product Gases	Hydrocarbon Gases	Reformer Feed	JP-8X	Hydro-treater 550+°F Product
Composition, Lb/Hr:									
H_2	--	--	--	1785	--	55	--	--	--
C_1	--	--	--	--	--	90	--	--	--
C_2	--	--	--	--	--	145	--	--	--
C_3	--	--	--	--	--	185	--	--	--
iC_4	--	--	--	--	--	20	--	--	--
nC_4	--	--	--	--	--	90	--	--	--
C_5	--	--	--	--	--	20	--	--	--
H_2S	--	--	--	--	160	--	--	--	--
NH_3	--	--	--	--	340	--	--	--	--
H_2O	950	950	--	--	3,785	--	--	--	--
Reformer Feed	--	--	--	--	--	--	6,345	--	--
JP-8X	--	--	--	--	--	--	--	24,545	--
Tar Oil	46,670	--	--	--	--	--	--	--	--
750-°F Tar Oil	--	36,920	--	--	--	--	--	--	--
Fuel Oil	--	--	9,750	--	--	--	--	--	3,875
Total Flow, Lb/Hr	47,620	37,870	9,750	1,785	4,285	605	6,345	24,545	3,875
B/D	3,182	2,584	598	--	--	60(FOE)	555	1,960	285

LPS/lkg/87863
9/9/87

TABLE III

PROPERTIES OF PRODUCTS FROM TAR OIL DISTILLATION
GREAT PLAINS CASE 5:
MAXIMUM JP-8X PRODUCTION

	750-°F Tar Oil	750+°F Tar Oil
Stream No.	2	3
Flow Rate, B/D	2,584	598
API	9.5	-4.7
Elemental Analysis, wt%:		
C	81.3	85.3
H	8.7	8.1
N	0.7	1.4
S	0.4	0.4
O (Difference)	8.9	4.8
Solids, wt%	0	3.9±2.9
Ash, wt%	0	0.5±0.4
Ramscarbon Residue, wt%	0	10
Water, wt%	2.5±1.0	0
TBP Distillation, °F		
IBP	272	557
5%	333	683
10%	363	714
20%	407	755
30%	445	785
50%	513	831
70%	587	859
80%	633	874
90%	690	890
95%	728	899
FBP	919	925

LPS/lkg/87863
9/9/87

TABLE IV

HYDROTREATER PRODUCT PROPERTIES
GREAT PLAINS CASE 5
MAXIMUM JP-8X PRODUCTION

	Reformer Feed	JP-8X	550+°F
Stream No.	7	8	9
Flow Rate, B/D	555	1,960	285
API	50	33	21
Composition, Vol%:			
Paraffins	14	9	20
Naphthenes	70	72	65
Aromatics	16	19	15
Octane: Research	72	--	--
Motor	68	--	--
RVP, psi	3	--	--
V/L = 20, °F	195	--	--
ASTM D-86 IBP, °F	165	273	439
5%	180	343	490
10%	194	353	501
30%	232	380	529
50%	254	404	554
70%	273	433	617
90%	307	476	706
95%	325	495	732
FBP	392	596	767

LPS/lkg/87863
9/9/87

TABLE V

HYDROTREATER OPERATING CONDITIONS AND UTILITIES
GREAT PLAINS CASE 5
MAXIMUM JP-8X PRODUCTION

Operating Conditions	
Reactor Type	Ebullated Bed
Catalyst	NiW
Feed Rate, B/D	3,200
Reactor Temperature, °F	700
Recycle Gas Rate, SCF/Bbl	6,000
Catalyst Replacement Rate, Lb/Bbl	0.18 (Approx. $6/Lb)
Estimated Utilities	
Fuel, MMBtu/Hr	0
Power, KW	2,600
Steam, MLb/Hr:	
600 psig	(650)
100 psig	250
Cooling Water, GPM	580
Process Water, GPM	150

LPS/lkg/87863
9/9/87

APPENDIX G

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 6
PROFITABLE JP-8X PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

W16246/bf

TABLE OF CONTENTS

1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to produce JP-8X type aviation turbine fuel and chemical byproducts to maximize profit from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- A portion of the Tar Oil byproduct stream (40604 #/hr, 2713 BPSD) and the Neutral Oil stream produced in the Phenol Processing (4936 #/hr, 312 BPSD) is charged to the hydrotreater (Area 100).

- The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 550^{o}F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (3618 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.

- The hydrotreater produces 6 streams:

  - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

  - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.

  - Unstabilized naphtha which is sent to the naphtha stabilizer. After stabilization, to control vapor pressure, the naphtha is sent to storage and gasoline blending.

  - JP-8X turbine fuel which is sent to storage.

  - 550^{o}F+ unconverted bottoms product which is sent storage and fuel.

  - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

- Approximately 900 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.

. Hydrogen make-up for both the Hydrotreater (Area 100) and the Naphtha Hydrotreater (Area 600) is supplied from a PSA Hydrogen Unit (Area 300). High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft^3. This H_2, CO & CH_4 rich gas is recompressed into the methanation unit of the SNG plant.

. The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).

. The distillation removes the material boiling below 160^{o}F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.

. The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.

. The naphtha hydrotreater produces 4 streams:

- High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.

- Naphtha which is stabilized to control vapor pressure, and the sent to the aromatics recovery unit (Area 700).

- A low pressure off gas which is sent to the Stretford unit in the SNG plant.

- Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.

The hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.

1.1 Overall Process Description - cont'd

- Five streams are produced in the ARU plant.
    - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.
    - A small process water stream which is sent to the waste treatment plant in the SNG Plant.
    - Three product streams Benzene, Toluene & Xylene which are sent to storage.
- The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).
- Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.
- The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.
- The acid tar produced is water washed and mixed with light oil and sent to fuel.
- The remaining cresol/xylenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).
- Streams produced in the phenol extraction unit are:
    - Phenol product sent to storage
    - Crude Cresylic Acid sent to distillation (Area 900) or storage.
    - Wash Water sent to Water Treatment in the SNG Plant.
    - Waste Water sent to the Phenosolvan unit in the SNG Plant.
    - Neutral Oil sent to storage and fuel for the SNG Plant boilers.
- The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.

1.1 Overall Process Description - cont'd

- Streams produced in the crude cresylic acid distallation unit are:
  - o-Cresol product which is sent to storage.
  - m,p-Cresol product which is sent to storage.
  - Xylenol product which is sent to storage.
  - A heavy distillate which is combined with neutral oil in Area 800.
  - A crude phenol stream which is recycled to the Area 800.
  - A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas 160°F-distillate, and 550°F heavy oil produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

936	BPSD of Crude Phenol
725	BPSD of Crude Naphtha
2713	BPSD of Tar Oil
3312	BPSD of #6 Fuel Oil
8.72	MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

1938	BPSD of JP-8X turbine fuel
522	BPSD of 300°F - Naphtha for gasoline blending
317	BPSD of Phenol
56	BPSD of o-Cresol
131	BPSD of m,p-Cresol
75	BPSD of Xylenols
844	BPSD of 550°F+ for Fuel
202	BPSD of 160°F - Distillate for Fuel

1.2 Overall Material Balance - cont'd

Products

46 BPSD of Gasoline Blending Stock
315 BPSD of Benzene
112 BPSD of Toluene
15 BPSD of Xylene
5.60 MMSCFD equivalent SNG product credit due to HDT, & PSA purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	3312 BPSD
SNG Equivalent of Syn Gas & Purge Gas	3.12 MM SCFD
Power	4810 kW
Cooling Water	5250 GPM (30^{o}F rise)
Process Water	24 GPM

In addition the process utilizes steam as summarized below which was debited against boiler requirements.

HP Steam Import	58,200 #/H
MP Steam Import	9,300 #/H
LP Steam Export	7,050 #/H
Condensate Return	67,500 #/H

Figure 1 Case 6: Profitable JP-8X

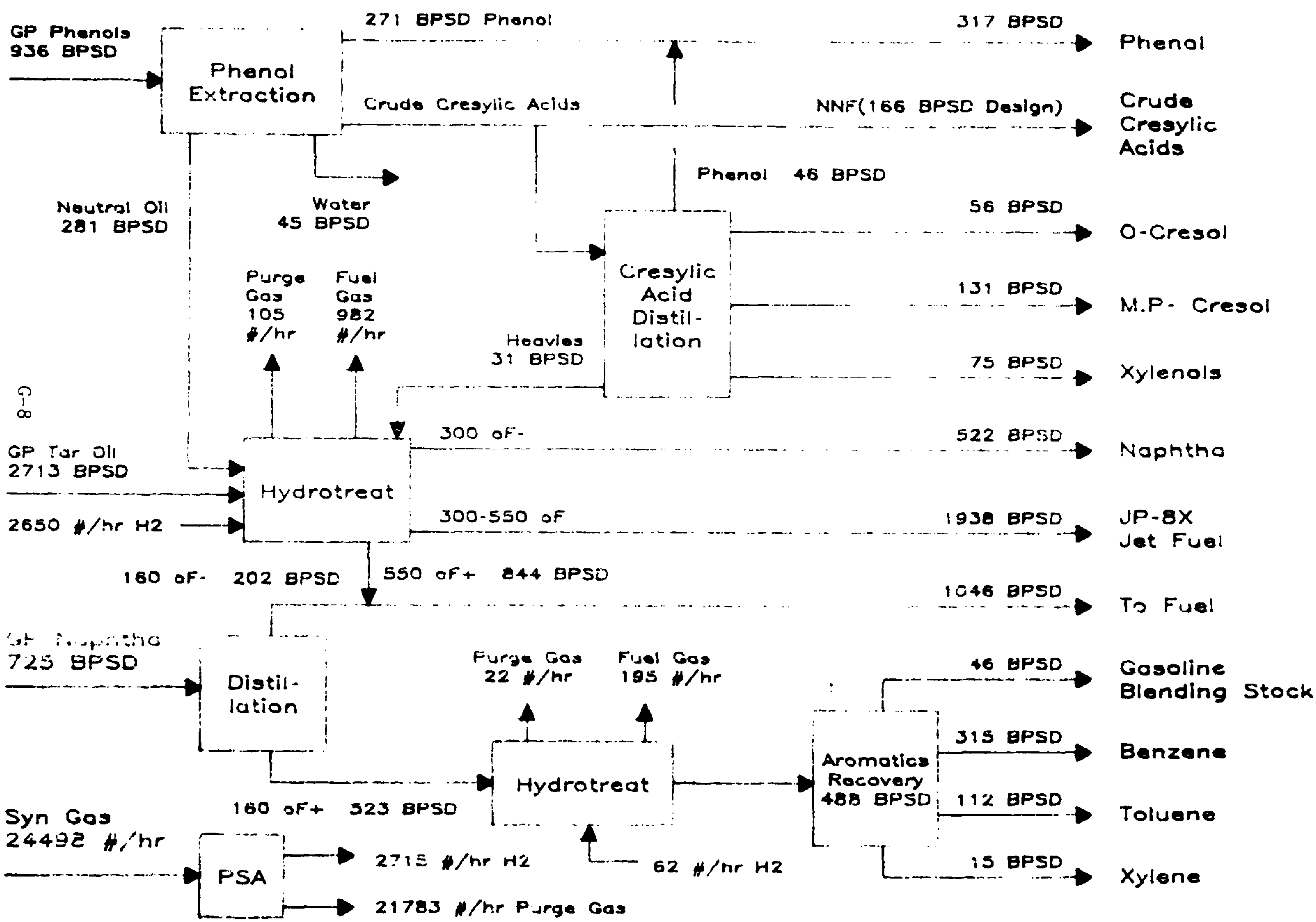


Table 1.1 Great Plains Case 6: Profitable JP8X Production

Tar Oil Feed=====>	40604	#/hr	2713	BPSD
Phenol Feed======>	14490	#/hr	936	BPSD
Crd Naphtha Feed=>	8738	#/hr	725	BPSD
Naphtha Product==>	5665	#/hr	522	BPSD
JP8X Product=====>	24318	#/hr	1938	BPSD
Phenol Product===>	4925	#/hr	317	BPSD
o-Cresol Prod====>	845	#/hr	56	BPSD
m,p-Cresol Prod==>	1974	#/hr	131	BPSD
Xylenols Prod====>	1070	#/hr	75	BPSD
Gasoline Stock===>	481	#/hr	46	BPSD
Benzene Prod=====>	4060	#/hr	315	BPSD
Toluene Prod=====>	1425	#/hr	112	BPSD
Xylene Prod======>	188	#/hr	15	BPSD
SNG Product Loss=>	5353	#/hr	3.1	MMSCFD
Fuel Oil Makeup==>	45852	#/hr	3312	BPSD

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
Oil Feeds					
Tar Oil	89.16	1.0268	40604		2713
Neut. Oil	9.76	1.0860	4445		281
Heavies	1.08	1.0800	491		31
Tot. Oil	100.00		45540		3025
H2	5.82		2650	1314.5	
Total	105.82		48190		
Products					
Purge Gas	0.23		105	32.6	
Fuel Gas	2.16		982	48.4	
Naphtha	12.69	0.7600	5781		522
JP-8X	53.40	0.8608	24318		1938
550 oF+	26.88	0.9950	12240		844
H2O in SW	8.96		4082	226.8	
H2S in SW	0.43		197	5.8	
NH3 in SW	1.06		485	28.5	
Total	105.82		48190		3304

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
HDT Nap	100.00	0.7600	5781		522
Stab Nap	98.00	0.7450	5665		522
Fuel Gas	2.00		116	2.7	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1564.3	460.9	36.6	401.4	7.8	4.7	2475.7
Prod. H2	1345.3	0.1	0.0	0.0	0.0	0.0	1345.4
Purge Gas	219.0	460.8	36.6	401.4	7.8	4.7	1130.3
#/hr							
Feed Gas	3154	12908	1612	6440	234	151	24498
Prod. H2	2712	3	0	0	0	0	2715
Purge Gas	442	12905	1612	6440	234	151	21783

Crude Naphtha Distillation	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

Aromatics Recovery

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	6154	488
Products				
Raffinate	7.82	0.7175	481	46
Benzene	65.97	0.8844	4060	315
Toluene	23.16	0.8718	1425	112
Xylene	3.05	0.8729	188	15
Total Products	100		6154	488

Phenol Extraction

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Cr. Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	982	48.4	17.7
Stabilizer FG	116	2.7	2.1
Naphtha Hdtr FG	195	10.8	3.5
Total Fuel Gas	1098	51.1	19.8

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	442	219.0	324	26.9
CO	12905	460.8	321	56.1
CO2	1612	36.6	0	0.0
C1	6440	401.4	1010	153.7
C2	234	7.8	1769	5.2
N2+Ar	151	4.7	0	0.0
Total	21783	1130.3	565	241.8

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Tar Oil	-40604	17000	-690.3			-2713
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
Fuel Gas	1098	18000	19.8	0.5	1020	
160 oF- distillate	2164	17400	37.7			202
550 oF+Hvy Oil	12240	18000	220.3			844
Import Steam	-60450	1000	-60.5			
Fuel Oil to Boiler	45779	18000	824.0			3307
Total	-63001		0.0	0.5		-21
Fuel Oil to Process Heaters	73	18000	1.3			5

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	8.72	59416
SNG Credit for PSA Purge gas	5.46	27127
SNG Credit for Hdtrs purge gas	0.14	947
Total SNG Production Loss	3.12	

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

For a description of the Hydrotreater process see Case 3 Section 2.1.

2.2 PSA Hydrogen Unit & Recompression (Area 300)

2.2.1 Hydrogen for both hydrotreaters will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig	
Temp.	65 °F	
Composition	mol%	
H2		63.19
CO		18.61
CO2		1.48
CH4		16.21
C2H6		0.31
COS, H2S, CS2		< 0.01
N2 + Ar		0.19
H2O		< 0.01

The PSA unit selectively absorbs all components expect H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

At the conditions given a 8 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

2.3 PSA Hydrogen Unit & Recompression (Area 300) - cont'd

2.3.1 Cont'd

The system uses 8 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-60301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

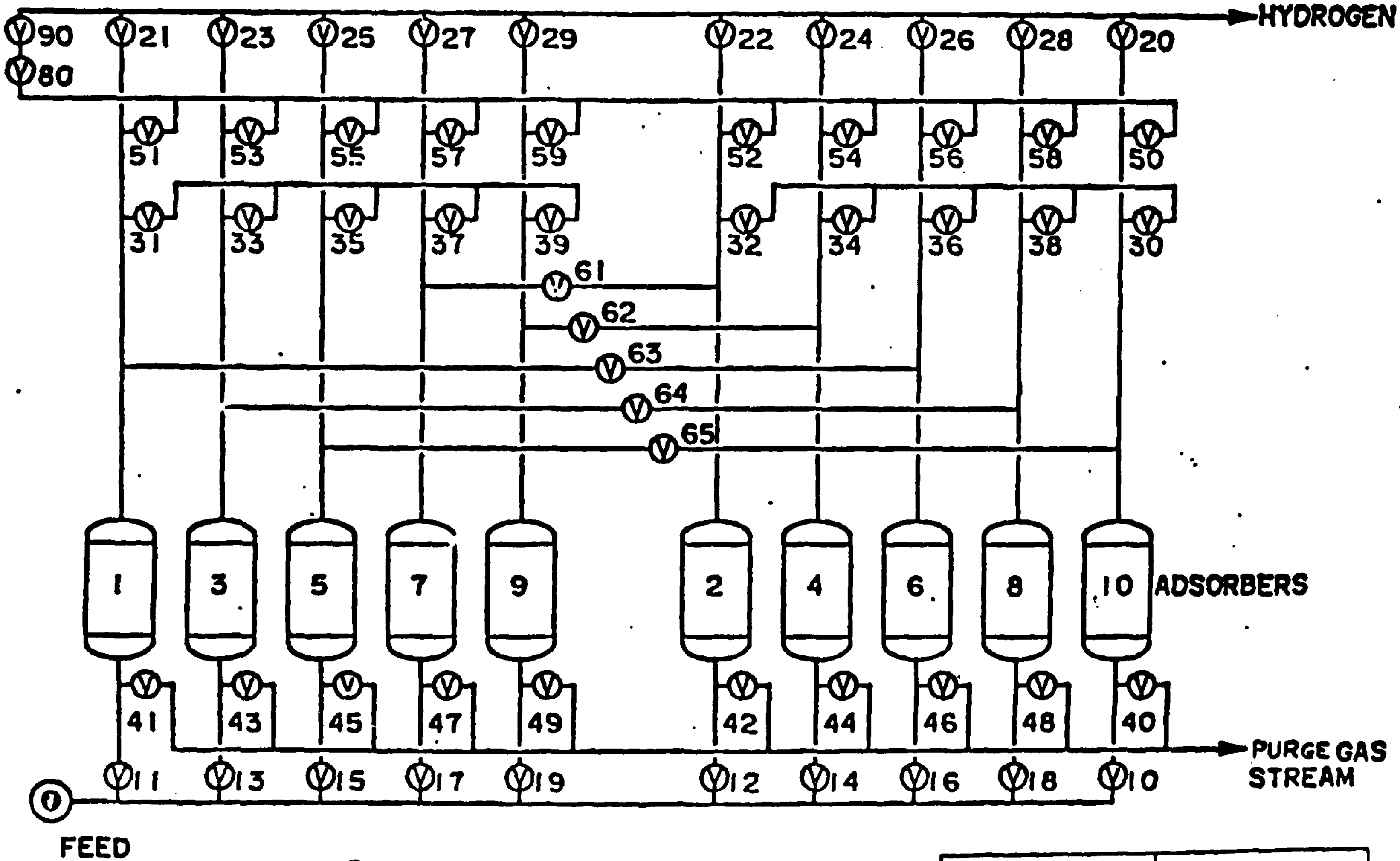
2.4 Phenol Stream (Area 800 and 900)

For a description of the Phenol Extraction (Area 800) and Cresylic Acid Distillation Units see Case 7 Section 2.1.

2.5 Naphtha Stream (Area 600 and 700)

For a description of the Naphtha Distillation and Hydrotreating Unit (Area 600) and the Aromatics Recovery Unit (Area 700) see Case 7 Section 2.2.

SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 8

△							LUMMUS — THE LUMMUS COMPANY
△	1/31/80	FOR SUBTASK 1.2	ML	E.S.			TITLE PSA HYDROGEN UNIT AREA 300; CLIENT AMOCO/DOE; PROJ NO 5571
REV	ISSUE DATE	DESCRIPTION	PROC ENGR	PROJ ENGR	APPR	APPR	CASE 6 — DWG NO - 5571-60301

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 6 - PROFITABLE JP-8X

AREA 100 - HYDROTREATER

TAG. NO.	DESCRIPTION
See Case 3 Area 100	

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

AREA 400 - STORAGE AREA

FB-401	Jet Fuel Storage Tank
FB-402	Naphtha Storage Tank
FB-403	Fuel Oil Storage Tank
FB-404	Blending Stock
FB-405	Benzene Storage
FB-406	Toluene Storage
FB-407	Xylene Storage
FB-409	Gasoline Storage
FB-411	Phenol Product Storage
FB-412	Crude Cresylic Acid Storage
FB-413	O-Cresol Storage
FB-414	M, P Cresol Stoage
FB-415	Xylenol Storage

3.0 CAPITAL COSTS

3.1 Equipment List - cont'd

CASE 6 - PROFITABLE JP-8X

TAG NO.		DESCRIPTION
AREA 400		STORAGE AREA
GA-401A/S		Tar/Tar Oil Feed Pump
GA-402A/S		Crude Phenol Feed Pump
GA-403A/S		Fuel Oil Transfer Pump
GA-404A/S		Naphtha Transfer Pump
GA-405A/S		Crude Naphtha Transfer Pump
GA-406A/S		Blending Stock Pump
GA-407A/S		Benzene Transfer Pump
GA-408A/S		Toluene Transfer Pump
GA-409A/S		Xylene Transfer Pump
GA-411A/S		Gasoline Transfer Pump
GA-413A/S		Crude Cresylic Acid Transfer Pump
GA-414A/S		O-Cresol Transfer Pump
GA-415A/S		M, P. Cresol Transfer Pump
GA-416A/S		Xylenol Transfer Pump
PA-401		Gasoline Blending Package
AREA 500	-	CATALYST HANDLING
See Case 3 Area 500		
AREA 600	-	NAPHTHA DISTILL. AND HDT
See Case 7 Area 600		
AREA 700	-	AROMATICS RECOVERY
See Case 7 Area 700		
AREA 800	-	PHENOL EXTRACTION
See Case 7 Area 800		
AREA 900	-	CRESYLIC ACID DISTILLATION
See Case 7 Area 900		

W16246/bf

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate for this case is a factored type estimate using the T.I.C. values developed for the various cases referenced in this project.

The total investment costs are scaled to the capacity requirement of ths case use a o.6 exponent.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs for Areas 100 thru 700 and 30% for Areas 800 & 900.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of $)

	Case 6
Area 100 Hydrotreater	$20,702
Area 300 PSA & Recompression	6,760
Area 400 OSBL	8,511
Area 500 Catalyst Handling	1,285
Area 600 Naph. Dist & HDT	4,615
Area 700 ARU	9,373
Area 800 Phenol Ext.	12,276
Area 900 Cresylic Acid Dist.	4,832
Subtotal	$68,354
Area 700 ARU Solvent Inv.	100
Total	$68,454

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

This unit has the same capacity as the 100 Area of Case 3. Therefore, T.I.C. = $20,702.

3.2.3 Estimate Breakdown - Cont'd

Area 300

This unit has a 71% capacity of the 300 Area of Case 3.
Therefore T.I.C. = $(0.71)^{0.6}$ (8300) = $6760.

Area 400

Tar Oil Stream Storage 75% Case 1
TIC = $(0.75)^{.6}$ (5 m) = $4200

Phenol Stream Storage = 100% Case 7
TIC = $3,016

Naphtha Stream Storage 100% Case 7
TIC = $3058
Subtotal $10,274

Less Duplicate Pipe & Rack -1763
Total = $8,511

Area 500

The capacity of this unit is identical to the 500 Area of Case 3. Therefore T.I.C. = $1,285

Area 600

This unit has a capacity of 100% of the 600 Area of Case 7.
Therefore T.I.C. = $4,615

Area 700

This unit has a capacity of 100% of the 700 Area of Case 7.
Therefore T.I.C. = $9,373

Area 800

This unit has a capacity of 100% of the 800 Area of Case 7.
Therefore T.I.C. = $12,276

Area 900

This unit has a capacity of 100% of the 900 Area of Case 7.
Therefore = T.I.C. = $4,852

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require men/shift to operate the plant broken down as follows:

Foreman	2
Control Room	2
HDT Operator	2
PSA & relief man	1
Naphtha Distil. & HDT	2
ARU	2
Phenol Ext.	1
Cresylic Acid Dist.	1
	13 Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 7 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	13 positions x 4 people/position -	52
Supervisor & Admin.		6
QC Technician		2
Maintenance		7
Other (Stores or Janitorial)		1
Total		68

4.2 Utilities

The following utilities have been estimated:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	3312 BPSD	$16/Bbl (a)	52992
SNG equivalent of Syn Gas & Purge Gas	3.12 MMSCFD	$3.80/MM Btu (b)	11619
Cooling Water	5250 GPM	$0.155/MGal (c)	1172
Power	4810 kW	$0.04/kWH (c)	4618
Process Water	24 GPM	$0.45/MGal (c)	15

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

4.2 Utilities - cont'd

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

Catalyst & Chem.	Use	Cost	$/SD
Nap. HDT Cal	0.021 #/Bbl	$3.00/#	33
HDT Cat.	0.30 #/Bbl	$3.00/#	2720
Inhibitors	50 PPM	$10/Gal	52
ARU Solvent	18 #/D	$2.10/#	50
H_2SO_4	7100	$0.04/#	285
			$3140

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

0.00005 x 68,354,000 = $3418/SD

W16246/bf

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-8X and by-product chemicals are proposed to be located to the east of the Phenosolvan Unit and Water Treatment Area of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 340' x 390' will be surrounded by an access road and will be divided by two central east-west roads. Areas 100, 300 & 500 will be located to the north, Areas 800 & 900 next and then Areas 600 & 700.

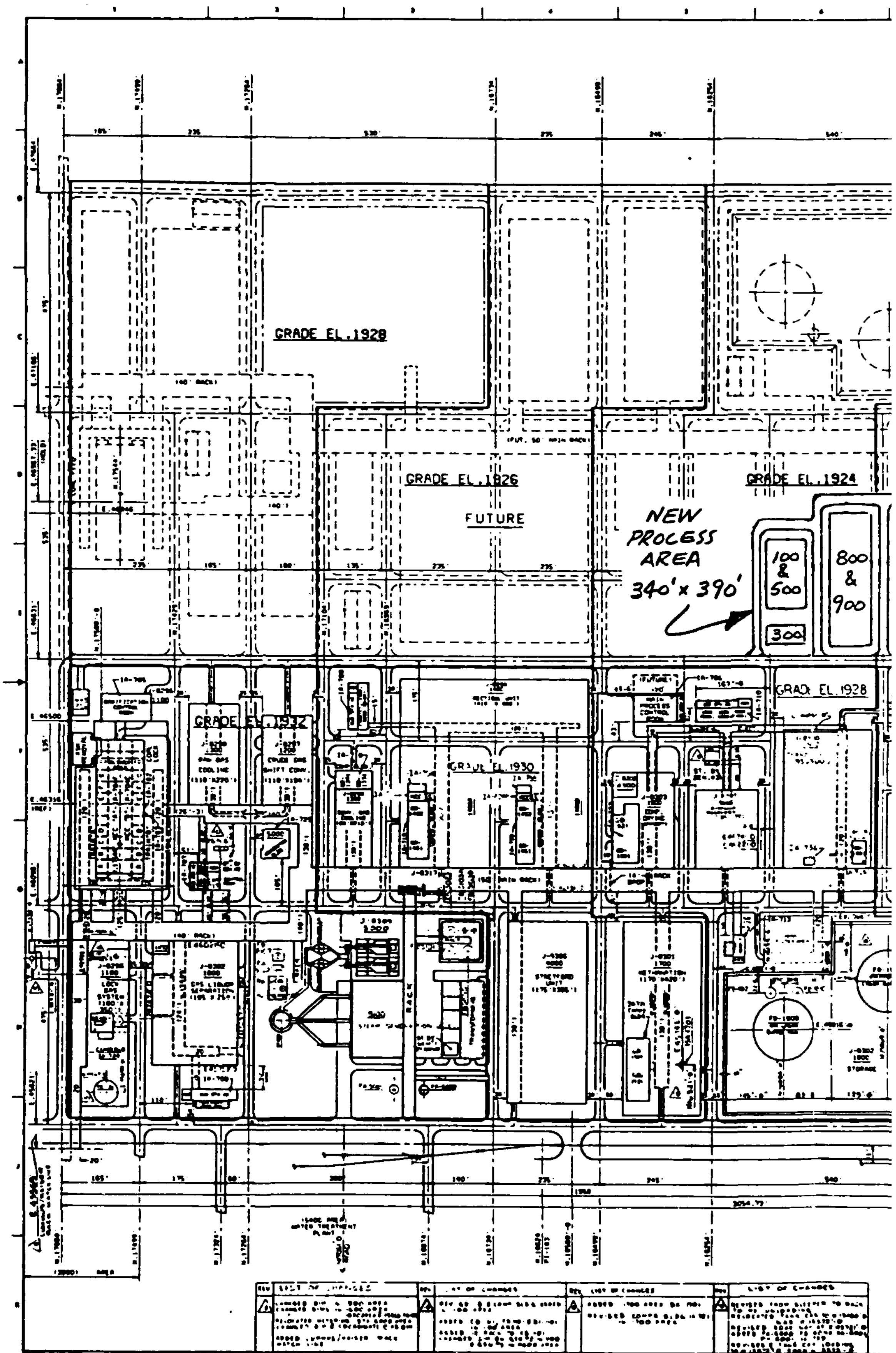
A diked storage tank area approx. 375' x 325' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

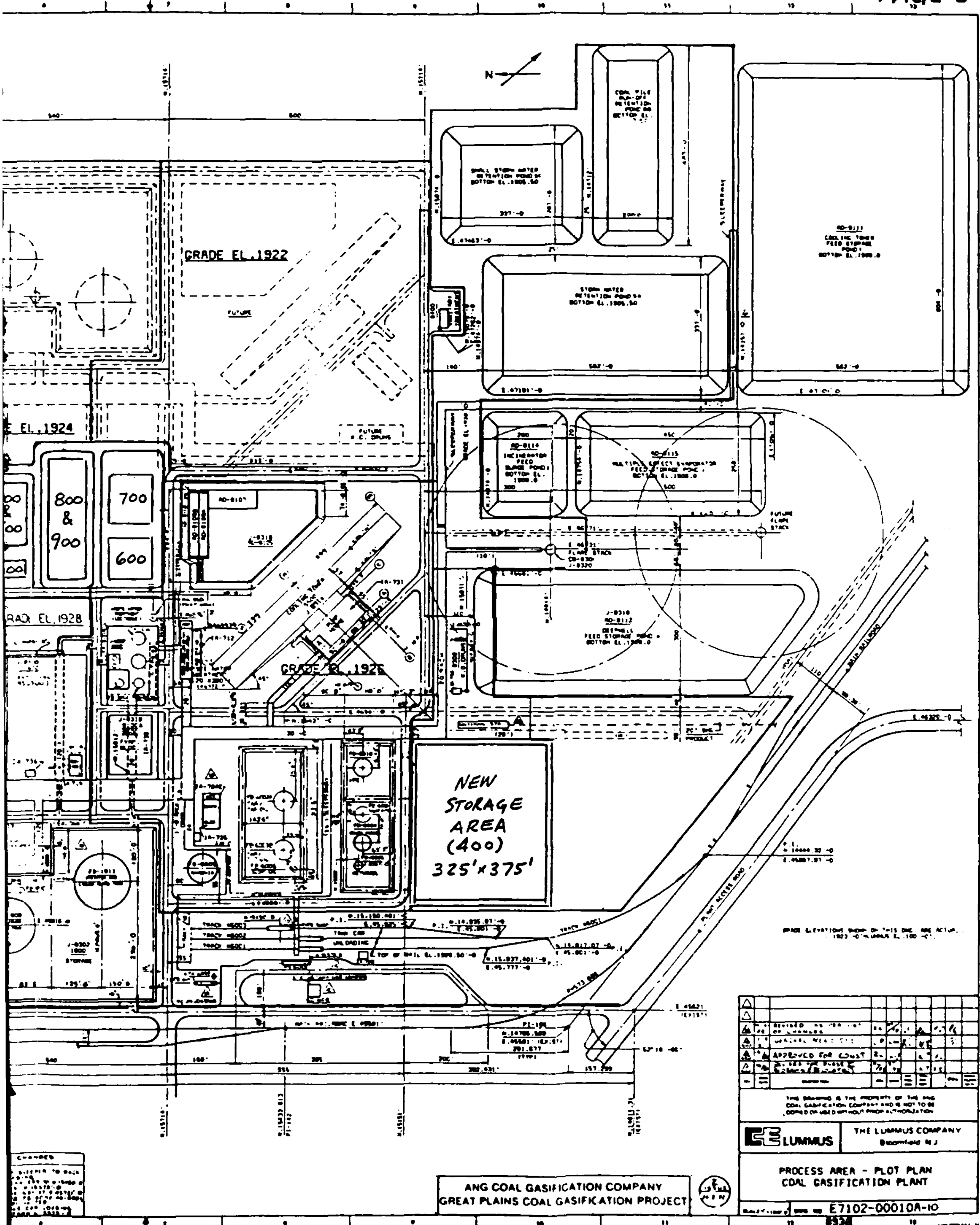
5.2 Unit Tie-Ins

Approximately 2500 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

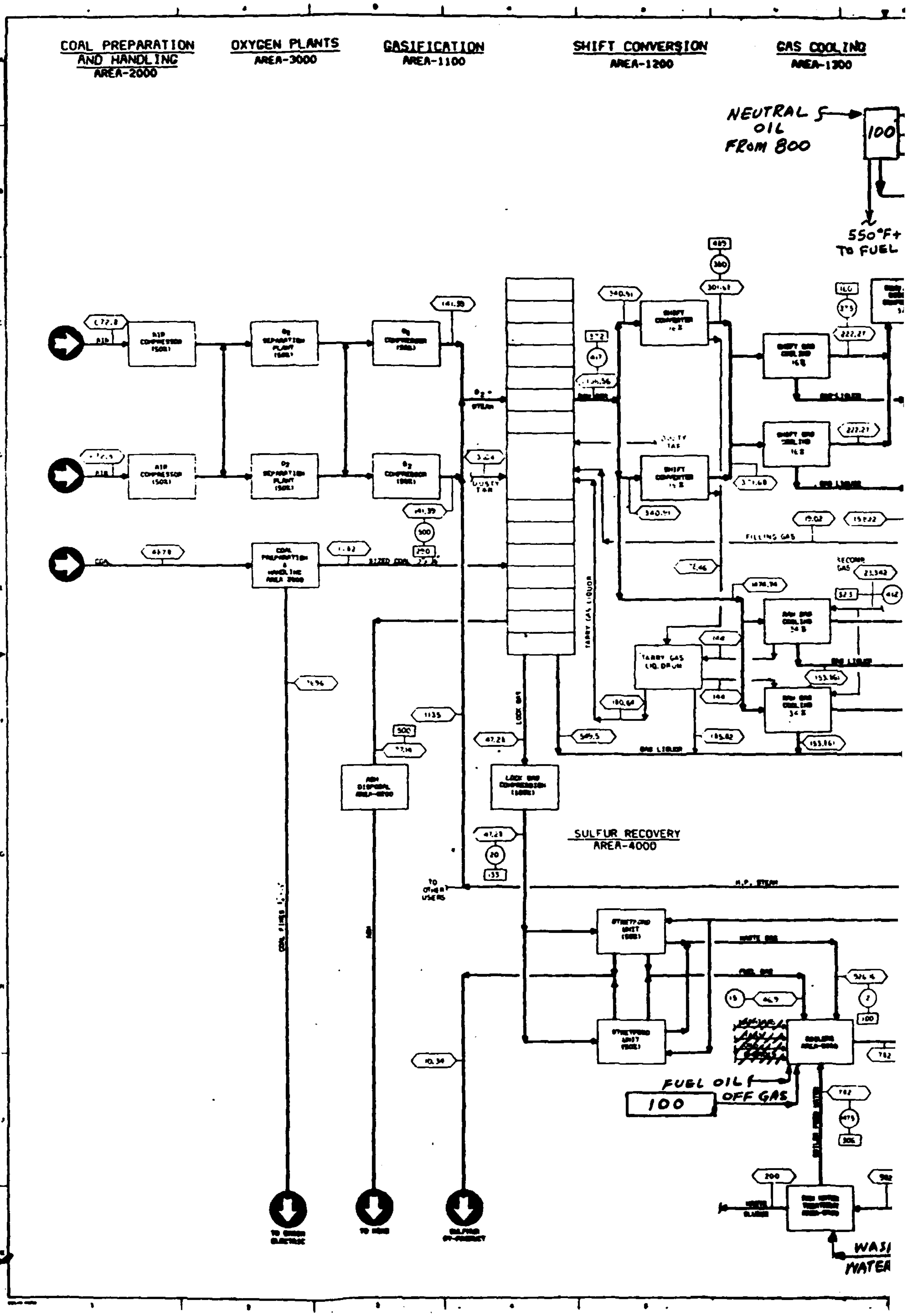
New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines has not been prepared for this case but will be similar to a combination of Cases 3 and 7 with the utility lines of like services being combined.

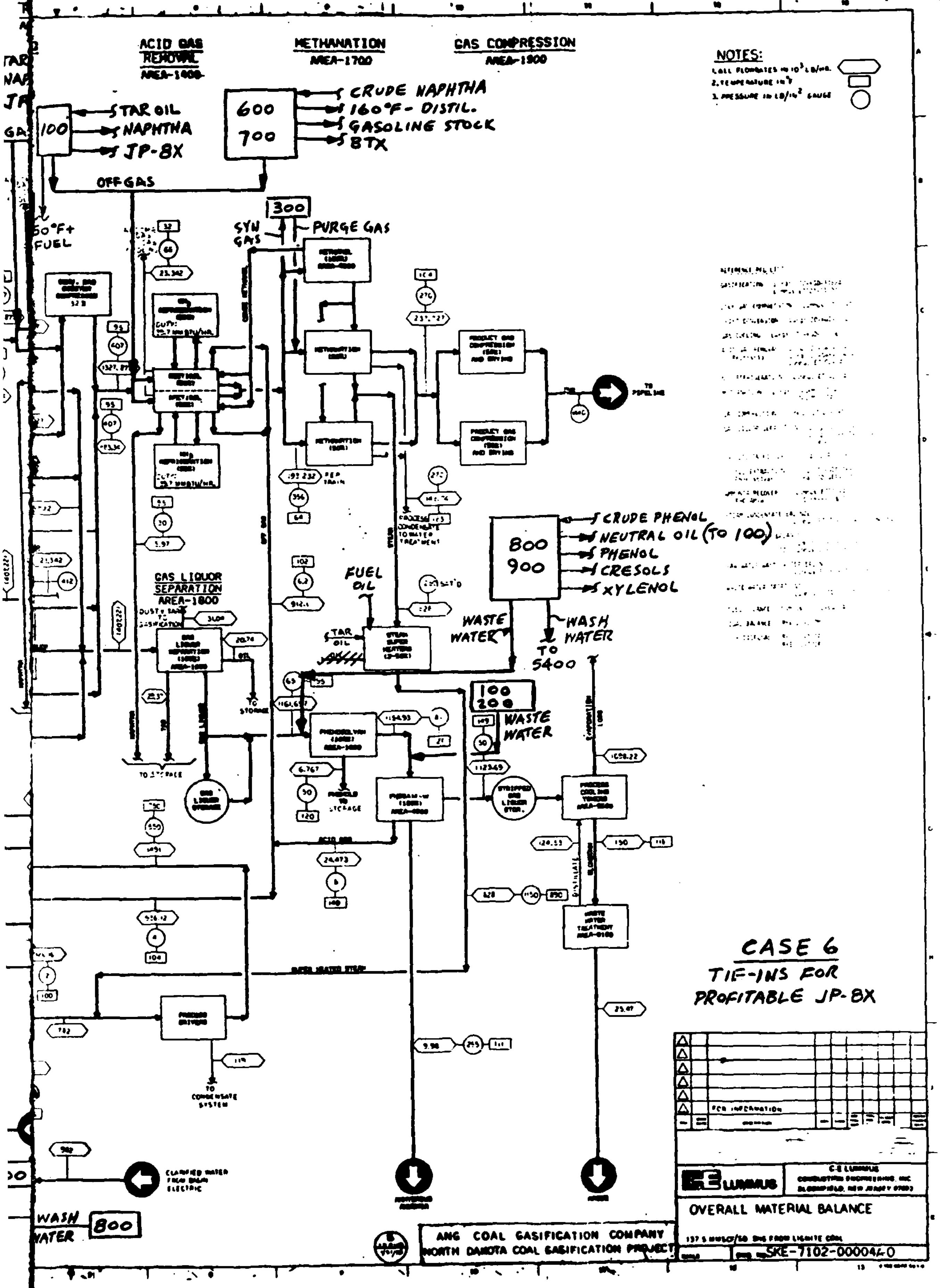

GRADE EL.1928
GRADE EL.1926
FUTURE
GRADE EL.1924
NEW
PROCESS
AREA
340'x 390'
100
&
500
800
&
900
300
GRADE EL.1932
GRADE EL.1930
GRADE EL.1928


N
GRADE EL.1922
FUTURE
FUTURE F.C. DRUMS
EL.1924
800 & 900
700
600
GRAD EL.1928
GRADE EL.1926
NEW STORAGE AREA (400) 325'x375'
PLANT ACCESS ROAD
APPROVED FOR CONST
LUMMUS
THE LUMMUS COMPANY
Bloomfield N.J.
PROCESS AREA - PLOT PLAN
COAL GASIFICATION PLANT
ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT
E7102-00010A-10

CASE 6



COAL PREPARATION AND HANDLING
AREA-2000
OXYGEN PLANTS
AREA-3000
GASIFICATION
AREA-1100
SHIFT CONVERSION
AREA-1200
GAS COOLING
AREA-1300
NEUTRAL OIL FROM 800
100
550°F+ TO FUEL
SULFUR RECOVERY
AREA-4000
FUEL OIL
OFF GAS
100
WASH WATER


ACID GAS REMOVAL
AREA-1400
METHANATION
AREA-1700
GAS COMPRESSION
AREA-1900
NOTES:
100
STAR OIL
NAPHTHA
JP-8X
600
700
CRUDE NAPHTHA
160°F - DISTIL.
GASOLINE STOCK
BTX
OFF GAS
300
SYN GAS
PURGE GAS
650°F+ FUEL
TO PIPELINE
800
900
CRUDE PHENOL
NEUTRAL OIL (TO 100)
PHENOL
CRESOLS
XYLENOL
FUEL OIL
WASTE WATER
WASH WATER
TO 5400
TAR OIL
GAS LIQUOR SEPARATION
AREA-1800
100
200
WASTE WATER
TO STORAGE
TO CONDENSATE SYSTEM
CLARIFIED WATER FROM BASIN ELECTRIC
WASH WATER
800
CASE 6
TIE-INS FOR
PROFITABLE JP-8X
LUMMUS
OVERALL MATERIAL BALANCE
ANG COAL GASIFICATION COMPANY
NORTH DAKOTA COAL GASIFICATION PROJECT
SKE-7102-000046-0

APPENDIX H

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 7
MAXIMUM PROFIT
SUBTASK 1.2
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE JAN 30, 1988

w16235/bf

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to maximize the profit from the processing of Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- Two by product streams are processed.
  - Phenol 14490 #/hr (936 BPSD)
  - Naphtha 8738 #/hr (725 BPSD)
- The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).
- The distillation removes the material boiling below 160^{o}F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.
- The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.
- The naphtha hydrotreater produces 4 streams:
  - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4.
  - Naphtha which is stabilized to control vapor pressure, and then sent to the aromatics recovery unit (Area 700).
  - A low pressure off gas which is sent to the Stretford unit in the SNG plant.
  - Wastewater containing, NH4OH and NH4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H2S and NH3.
- Hydrogen make-up for the Hydrotreater is supplied from a PSA Hydrogen Unit. High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining gas is available at low pressure (5 psig) and has a fuel value of about 565 BTU/ft^3. This gas is sent to the main boilers in the SNG plant.

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- The hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.

- Five streams are produced in the ARU plant.

  - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.

  - A small process water stream which is sent to the waste treatment plant in the SNG Plant.

  - Three product streams Benzene, Toluene & Xylene which are sent to storage.

- The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).

- Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.

- The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.

- The acid tar produced is water washed and mixed with light oil and sent to fuel.

- The remaining cresol/xylenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).

- Streams produced in the phenol extraction unit are:

  - Phenol product sent to storage

  - Crude Cresylic Acid sent to distillation (Area 900) or storage.

  - Wash Water sent to Water Treatment in the SNG Plant.

1.1 Overall Process Description - cont'd

- Waste Water sent to the Phenosolvan unit in the SNG Plant.

- Neutral Oil sent to storage and fuel for the SNG Plant boilers.

. The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.

. Streams produced in the crude cresylic acid distallation unit are:

- o-Cresol product which is sent to storage.

- m,p-Cresol product which is sent to storage.

- Xylenol product which is sent to storage.

- A heavy distillate which is combined with neutral oil in Area 800.

- A crude phenol stream which is recycled to the Area 800.

- A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. Detailed material balances for the Phenol Extraction & Cresylic Acid Distillation can be found in Appendix A and for the Naphtha Distillation & HDT unit Appendix B. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas, neutral oil and 160°F minus distillate produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

725	BPSD of Naphtha Feed
936	BPSD Phenol Feed
	BPSD of #6 Fuel Oil
0.20	MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

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1.2 Overall Material Balance - cont'd

Products

317 BPSD of Phenol
56 BPSD of o-Cresol
131 BPSD of m,p-Cresol
75 BPSD of Xylenols
312 BPSD of Neutral Oil for Fuel
202 BPSD of 160°F - Distillate for Fuel
46 BPSD of Gasoline Blending Stock
315 BPSD of Benzene
112 BPSD of Toluene
15 BPSD of Xylene

0.03 MMSCFD Equivalent SNG product credit due to HDT purge gas return to SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	1193 BPSD
SNG Equivalent of Syn Gas	0.17 MM SCFD
Power	525 kW
Cooling Water	4000 GPM (30°F rise)
Process Water	3 GPM

In addition the process imports/exports steam and returns condensate as follows:

HP Steam	58,200 #/HR Import
MP Steam	15,900 #/HR Import
LP Steam	6,900 #/HR Export
Condensate return	74,100 #/HR

Figure 1 Case 7:Maximum Profit

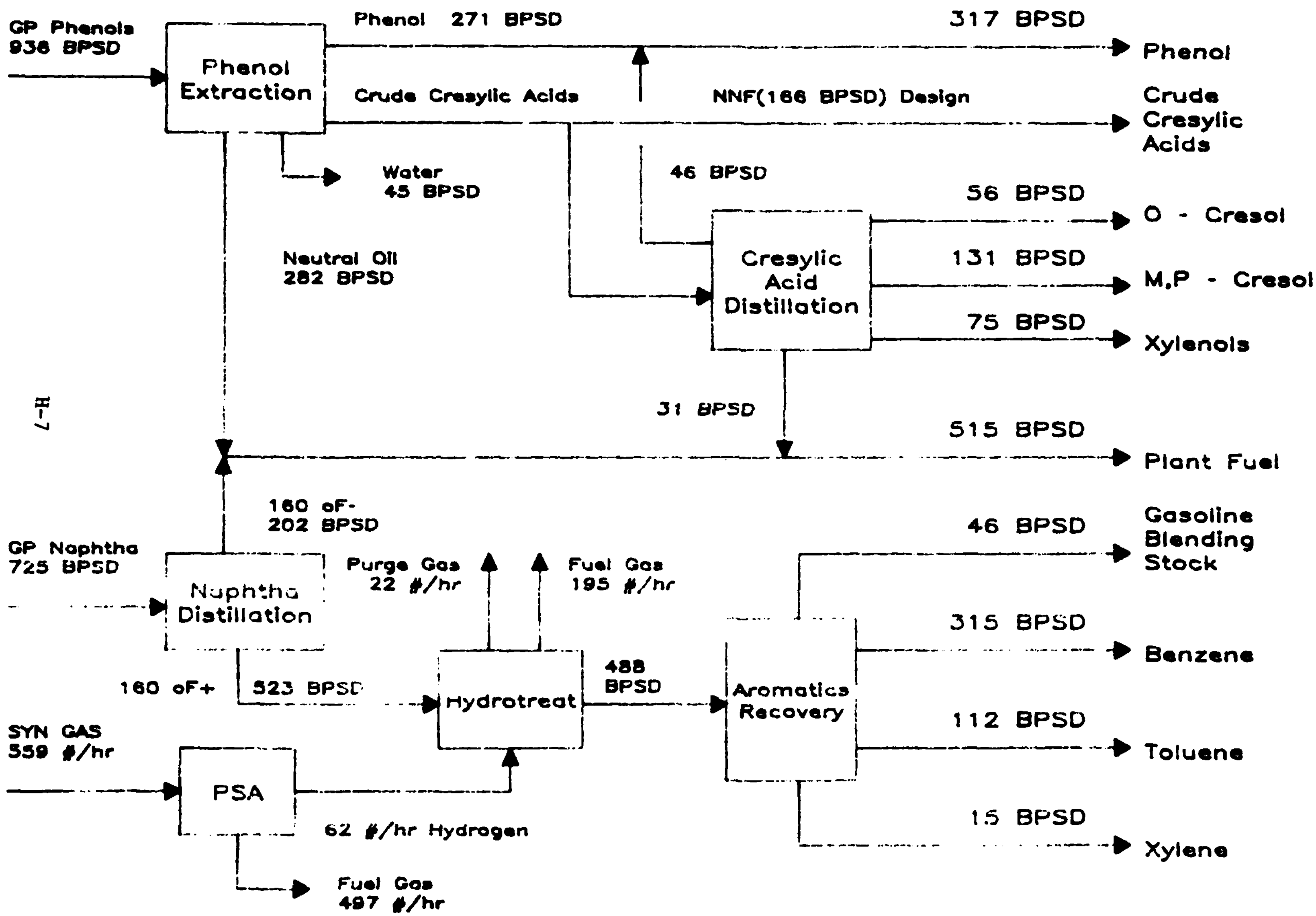


Table 1.1 Great Plains Case 7:Maximun Profit

Phenol Feed======>	14490	#/hr	936	BPSD
Crd Naphtha Feed=>	8738	#/hr	725	BPSD
Phenol Product===>	4925	#/hr	317	BPSD
o-Cresol Prod====>	845	#/hr	56	BPSD
m,p-Cresol Prod==>	1974	#/hr	131	BPSD
Xylenols Prod====>	1070	#/hr	75	BPSD
Gasoline Stock===>	481	#/hr	46	BPSD
Benzene Prod=====>	4060	#/hr	315	BPSD
Toluene Prod=====>	1425	#/hr	112	BPSD
Xylene Prod======>	188	#/hr	15	BPSD
SNG Product Loss=>	300	#/hr	0.17	MMSCFD
Fuel Oil Makeup==>	16517	#/hr	1193	BPSD

Crude Naphtha

Distillation	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	35.8	10.5	0.8	9.2	0.2	0.1	56.6
Prod. H2	30.8	0.0	0.0	0.0	0.0	0.0	30.8
Purge Gas	5.0	10.5	0.8	9.2	0.2	0.1	25.9
#/hr							
Feed Gas	72	295	37	147	5	3	560
Prod. H2	62	0	0	0	0	0	62
Purge Gas	10	295	37	147	5	3	498

Aromatics Recovery

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	6154	488
Products				
Raffinate	7.82	0.7175	481	46
Benzene	65.97	0.8840	4060	315
Toluene	23.15	0.8715	1425	112
Xylene	3.05	0.8729	188	15
Total Products	100		6154	488

Phenol Extraction

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Feed Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating

Component	#/hr	#Mol/hr	MMBTU/hr
Naphtha Hdtr FG	195	10.8	3.7

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	10	5.0	324	0.6
CO	295	10.5	321	1.3
CO2	37	0.8	0	0.0
C1	147	9.2	1010	3.5
C2	5	0.2	1769	0.1
N2+Ar	3	0.1	0	0.0
Total	498	25.9	565	5.5

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
PSA Purge Gas	498	11102	5.5	0.24	565	
Fuel Gas	195	19000	3.7	0.10		
160 oF- distillate	2164	17400	37.7			202
Neutral Oil	4936	15000	74.0			312
Import Steam	-67200	1000	-67.2			
Fuel Oil to Boiler	16517	18000	297.3			1193
Total	-66118		0.0	0.3		46

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	0.20	1359
SNG Credit for Hdtrs purge gas	0.02	164
Total SNG Production Loss	0.17	

2.0 PROCESS DESCRIPTION

2.1 Phenol Stream

2.1.1 Phenol Extraction

Data for the design of the Phenol Extraction Unit (Area 800) were provided to Lummus by ANG. The basic processing step used in this unit is a dual solvent extraction to recover the phenol product. Referring to drawing D5571-70801A and B and the material balance (Appendix A) the flow is as follows:

- Crude Phenol from the Great Plains Plant is charged to the Crude Phenol Column DA-801 from Surge Drum FA-801 through Feed Pump GA-801. Recycled phenol streams from Cresylic Acid Distillation (Area 900) and the Phenol Column DA-804 overheads are also charged to column DA-801. The bottoms from DA-801 is pumped by GA-802 to Acid Flash Column DA-802. The overhead from DA-802 contains light ends and phenol. Non condensibles are relieved to flare. The overhead liquids are pumped by GA-803 and GA-804 to the light ends column DA-803. Water condensed in the overheads is separated and sent to the Phenosolvan Unit in the SNG Plant.

- The Crude Phenol Column overhead enters the Light Ends Column DA-803. The overhead light ends product is sent to SNG Plant Fuel. Water condensed in the overhead drum is sent to the Phenosolvan Unit in the SNG Plant. The bottoms from the Light Ends Column are pumped to Phenol Column via Light Ends Bottoms Pump GA-809.

- Bottoms from the Light Ends Column are distilled in the Phenol Column DA-804 to produce an overhead stream which is returned to the Crude Phenol Column, DA-801 a side draw stream of 99.8% pure phenol product, and a bottoms stream which is pumped to Acid Flash Column DA-802 via Phenol Column Bottoms Pump GA-813. A small amount of sulfuric acid is added to the tower in order to control product purity.

- The Phenol Product is drawn off DA-804 to Phenol Draw-off Pot FA-804 and pumped by Phenol Drawoff Pump GA-814 through Phenol Product Cooler EA-810 to Phenol Product Day Tank FB-803. From the day tank, phenol product is pumped to storage via GA-815, Phenol Product Pump.

2.0 PROCESS DESCRIPTION

2.1.1 Phenol Extraction - cont'd

- The bottom streams from DA-801, Crude Phenol Column, and DA-804 Phenol Column are sent to Acid Flash Column DA-802. This combined stream contains the cresylic acid, neutral oil and heavies. This material is flashed over sulfuric acid to remove pyridine type substances. The overhead product of the acid flash is pumped to Extraction Column DA-805. The bottoms product is an acid tar, and is water washed in FD-801 & FD-802, 1st and 2nd stage Water Wash Tanks, to remove acid materials, and then routed to the Great Plains Fuel pool together with other fuel streams recovered in the Phenol Extraction and Cresylic Acid Distillation Units, (Areas 800 and 900). The combined stream is called "neutral oil."

- The acid flash overhead from DA-804 is extracted with hexane and methanol/water in Extractor Column DA-805. Hexane enters the extractor column at the bottom and preferentially absorbs the oil components. The hexane/oil mixture exits the top of DA-805.

- Methanol/water solution enters the top of the Extractor Column. The methanol/water preferentially adsorbs the phenolic compounds. Methanol/water/phenolic mixture exits the bottom of DA-805.

- The oil components are stripped from the hexane in the Hexane Column DA-806. The hexane is recycled to Extractor Column DA-804. The oil is pumped by Hexane Column Bottoms Pump GA-822 through Neutral Oil Cooler EA-815 to Neutral Oil Day Tank FB-802. From FB-802 Neutral Oil can be pumped to fuel or storage by GA-812 Neutral Oil Pump.

- Make-up Hexane is added as needed from Hexane Storage Tank FB-804 by GA-824 Hexane Make-up Pump.

2.0 PROCESS DESCRIPTION

2.1.1 Phenol Extraction - cont'd

- The phenolics are recovered from the methanol/water solution in Methanol Column DA-807. The methanol/water is condensed overhead and recycled to the extractor column by Methanol Column Reflux Pump GA-826. The phenolics are pumped to Drying Column DA-808 by Methanol Column Bottoms Pump GA-825 through Methanol Column Bottoms Cooler EA-818.

- Drying Column DA-808 is reboiled to remove water carry-over from the phenolic product. The dry Crude Cresylic Acid leaves the bottom of DA-808. Product is pumped to either Cresylic Acid Distallation (Area 900) or through the Crude Cresylic Acid Cooler EA-821 to the Cresylic Acid Day Tank, FB-805 by Drying Column Bottoms Pump GA-828.

- Crude Cresylic Acid from Cresylic Acid Day Tank is pumped to storage by Crude Cresylic Acid Pump GA-829.

2.1.2 Cresylic Acid Distillation

Data for the design of the Cresylic Acid Distillation Unit (Area 900) were provided to Lummus by ANG. The basic process used is a series of distillation columns to recover progressively higher boiling products. Referring to drawing D5571-70901 and the material balance (Appendix A) the flow is as follows:

- Dry Crude Cresylic Acid from Phenol Extraction (Area 800) is chargeed to O-Cresol Column DA-901.

- o-Cresol Reboiler EA-901 uses HP Steam to reboil the column. Liquid distillate product is returned to the Crude Phenol Column in Area 800. o-Cresol is a side draw product. Column DA-901 bottoms are feed to m,p-Cresol Column DA-902.

- o-Cresol product is stripped in o-Cresol stripper, DA-902, o-Cresol product is pumped by o-Cresol Stripper Pump GA-902 through o-Cresol Product Cooler EA-903 to o-Cresol Day Tank FB-901, o-Cresol product is pumped to storage by o-Cresol Product Pump GA-910.

2.0 PROCESS DESCRIPTION

2.1.2 Phenol Extraction - cont'd

- o-Cresol column bottoms is charged to the m,p-Cresol Column.

- m,p-Cresol product is recovered overhead. m,p-Cresol is pumped from m,p-Cresol Reflux Drum FA-902 by m,p-Cresol Reflux Pump GA-906 through m,p-Cresol Product Cooler EA-908 to the day tank FB-902. m,p-Cresol product is pumped to storage by the product pump GA-911.

- m,p-Cresol column bottoms is feed to DA-905 xylenol column. The xylenol product is recovered overhead. Xylenol is pumped from Xylenol Reflux Drum FA-903 by Xylene Reflux Pump GA-908 through Xylenol Product Cooler EA-911 to Xylenol Day Tank FB-903. Xylenol product is pumped to storage by Xylenol Product Pump GA-912.

- Xylenol column bottoms contains the undesirable heavies. Xylenol column bottoms are pumped by Xylenol Bottoms Pump GA-907 through Cresylic Acid By-product Cooler EA-912 to SNG plant fuel via the Neutral Oil system in Area 800.

2.2 Naphtha Stream

2.2.1 Naphtha Distillation & Hydrotreating (Area 600)

Operating conditions for the naphtha distillation and hydrotreater were provided to Lummus by Amoco and these conditions are presented in Table 2.1. The basic processing steps selected were a distillation to produce a 160^{o}F+ feed stock and a fixed bed hydrotreater. Referring to drawing D5571-70601 and the material balance (Appendix B) the flow is as follows:

- The crude naphtha is charged to the Naphtha Distillation Column DA-601 via Surge Drum FA-601 and Feed Pump GA-601.

- The column is reboiled with steam in EA-601 to produce a 160^{o}F+ bottoms product.

- The 160^{o}F- overheads are condensed in EA-602 and sent to fuel via GA-603.

- The 160^{o}F+ Distillation Column bottoms is charged to the HDT surge drum FA-603 via GA-602.

2.0 PROCESS DESCRIPTION

2.2.1 Hydrotreater - cont'd

- $160^{o}F+$ naphtha is charged into the hydrotreater from surge tank FA-603 by charge pumps GA-604 through Feed/Effluent Exchanger EA-603.

- The charge oil is combined with feed hydrogen gas from heater EA-604 prior to entering the feed/effluent exchanger. The preheated mixture is then charged to the reactor DC-601.

- The reactor DC-601A operates adiabatically with an average bed temperature of $450^{o}F$.

- The effluent from DC-601 is cooled in EA-603 and flows through exchangers EA-605 and EA-606. Process water is injected prior to EA-606 to convert the H2S and NH3 in the gas to an aqueous NH4OH/NH4HS solution.

- The cooled mixture then passes into the High Pressure/Low Temperature Separator FA-605 where hydrogen rich gas leaves overhead. A portion of this high pressure gas is purged to remove H_2S and light gases from the loop and sent to the Rectisol Unit 1400 in the SNG plant to recover the hydrogen in the purge gas. The remaining gas is recirculated to reactor DC-601.

- The water phase from separator FA-606 goes to the PHOSAM Unit in the SNG plant to recover the H2S and NH3.

- The hydrocarbon phase from separator FA-606 is preheated in exchanger EA-605 and charged to the HDT Naphtha Stabilizer DA-602.

- The unstabilized naphtha is charged into DA-602 which is reboiled by MP Steam to stabilize the naphtha.

- Offgas from the Naphtha Stabilizer is sent to the SNG plant for fuel.

- The stabilized naphtha is cooled and sent to the aromatics recovery unit (Area 700).

2.0 PROCESS DESCRIPTION

2.2.1 Hydrotreater - cont'd

Table 2.1 Hydrotreater Conditions

Case 7 Maximum Profit	
Feed Stock	$160^{\circ}F$+ Naphtha
Reactor Type	Fixed Bed
Number of Stages	1
LHSV Hr^{-1}	1.0
Average Reactor Temperature	$450^{\circ}F$
Reactor Pressure	500 psig H_2 Partial Pressure
H2 Recycle Rate	2500 SCF/BBL
Catalyst	Ni-Mo
Catalyst Replacement	2 years @ $3/#

2.2.2 Aromatics Recovery Unit (Area 700)

This unit is based on the Shell Sulfolane Process licensed by Universal Oil Products. Referring to Drawings D5571-70701A and B the flow is as follows:-

Stabilized Naphtha from Naphtha Hydrotreater (Area 600) is charged to the Extraction Column DA-701 through Feed Surge Drum FA-701 by Feed Charge Pump GA-701. Lean Solvent is charged to the top of column DA-701. As the feed flows through the column, aromatic components are selectively dissolved in the solvent. Raffinate with very low aromatics content is withdrawin from the top of DA-701.

Rich solvent leaves the bottom of the extractor. After heat exchange in Lean/Rich Solvent Exchanger EA-702, the rich solvent is charged to the top of DA-703, Stripper.

The raffinate stream from the Extractor Column DA-701 overheads is cooled in raffinate cooler EA-701 and then contacted with wash water in Water Wash Column DA-702. Water removes any dissolved solvent from the raffinate. Raffinate leaving DA-702 overhead is pumped to Gasoline Blending Stock Storage. The solvent rich water from DA-702 flows to DA-705, Water Stripper.

Solvent accumulates in the bottom of Water Stripper DA-705 and is pumped back to the Recovery Column by Water Stripper Bottoms Pump GA-710. The rich water is returned to the Recovery Column as stripping steam generated via the Water Stripper Reboiler EA-709 by exchange with the hot circulating lean solvent.

2.2.2 Aromatics Recovery Unit (Area 700) - cont'd

A solvent regeneration system is included to guard against excessive solvent degradation. In normal operation a slipstream of solvent is routed to the Solvent Regenerator DA-706. Degraded solvent is periodically withdrawn from the bottom of DA-706.

In the Stripper, non-aromatic hydrocarbons, which are more volatile, are stripped from the solvent, removed overhead, condensed and recycled to the Extractor Column DA-701 for recovery.

The stripper bottoms consists of aromatics in the solvent. This stream is pumped to the Recovery Column DA-704 by Stripper Bottoms Pump GA-704.

In the Recovery Column DA-704, the aromatics are stripped from the solvent. Lean solvent leaves the column bottom and is returned to Extraction Column DA-701 by GA-707 Lean Solvent Pump after heat Exchange in Water Stripper Reboiler EA-709 and Lean/Rich Solvent Exchanger EA-702.

The aromatic product recovered overhead from the Recovery Column in fractionated to recover benzene, toluene and xylene product streams.

The recovery column overhead is pumped by Recovery Column Overhead Pump GA-709 to Clay Tower Surge Tank, FB-703. From FB-703 the aromatic stream is pumped by Clay Tower Feed Pump GA-715 through Clay Tower Feed/Effluent Exchanger EA-712 and Clay Tower Feed Heater EA-713 and then into Clay Towers DA-707A/B. In the Clay Tower, trace amounts of unsaturates and residual non-hydrocarbon impurities are removed.

After heat exchange in the Clay Tower Feed/Effluent Exchanger, the extract flows to benzene column DA-708. Benzene product is withdrawn from a tray near the top of the tower. After cooling in Benzene Product Cooler EA-715, benzene flows to Benzene Day Tank FB-704. Product from FB-704 is pumped to product storage by Benzene Product Pump GA-719.

Any water that accumulates in Benzene Column Reflux Drum FA-708 is permed to Waste Treatment by Benzene Column Water Pump GA-718.

2.2.2 Aromatics Recovery Unit (Area 700) - cont'd

Benzene column bottoms are pumped by Benzene Column Bottoms Pump GA-716 to Toluene column DA-709. The Toluene Product leaves overhead. Toluene is pumped from Toluene Column Reflux Drum FA-709 by Toluene Column Reflux Pump GA-721 through Toluene Product cooler EA-720 to Toluene Day Tanks FB-706A/B. Toluene from FB-706A/B is pumped to storage by Toluene Product Pump GA-723.

Xylene is taken as bottoms product from Toluene Column DA-709. Xylene is pumped by Toluene Column Bottoms pump GA-720 through Xylene Product Cooler EA-718 to Xylene Day Tank FB-705. Xylene from FB-705 is pumped to storage by Xylene Product Pump GA-722.

2.2.3 PSA Hydrogen Unit (Area 300)

Hydrogen for the naphtha hydrotreater will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig	
Temp.	65 ^{o}F	
Composition	mol%	
H2		63.19
CO		18.61
CO2		1.48
CH4		16.21
C2H6		0.31
COS, H2S, CS2		< 0.01
N2 + Ar		0.19
H2O		< 0.01

The PSA unit selectively absorbs all components expect H2 and produces a 99.99% vol. purity stream at about 345 psig and 80^{o}F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100^{o}F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

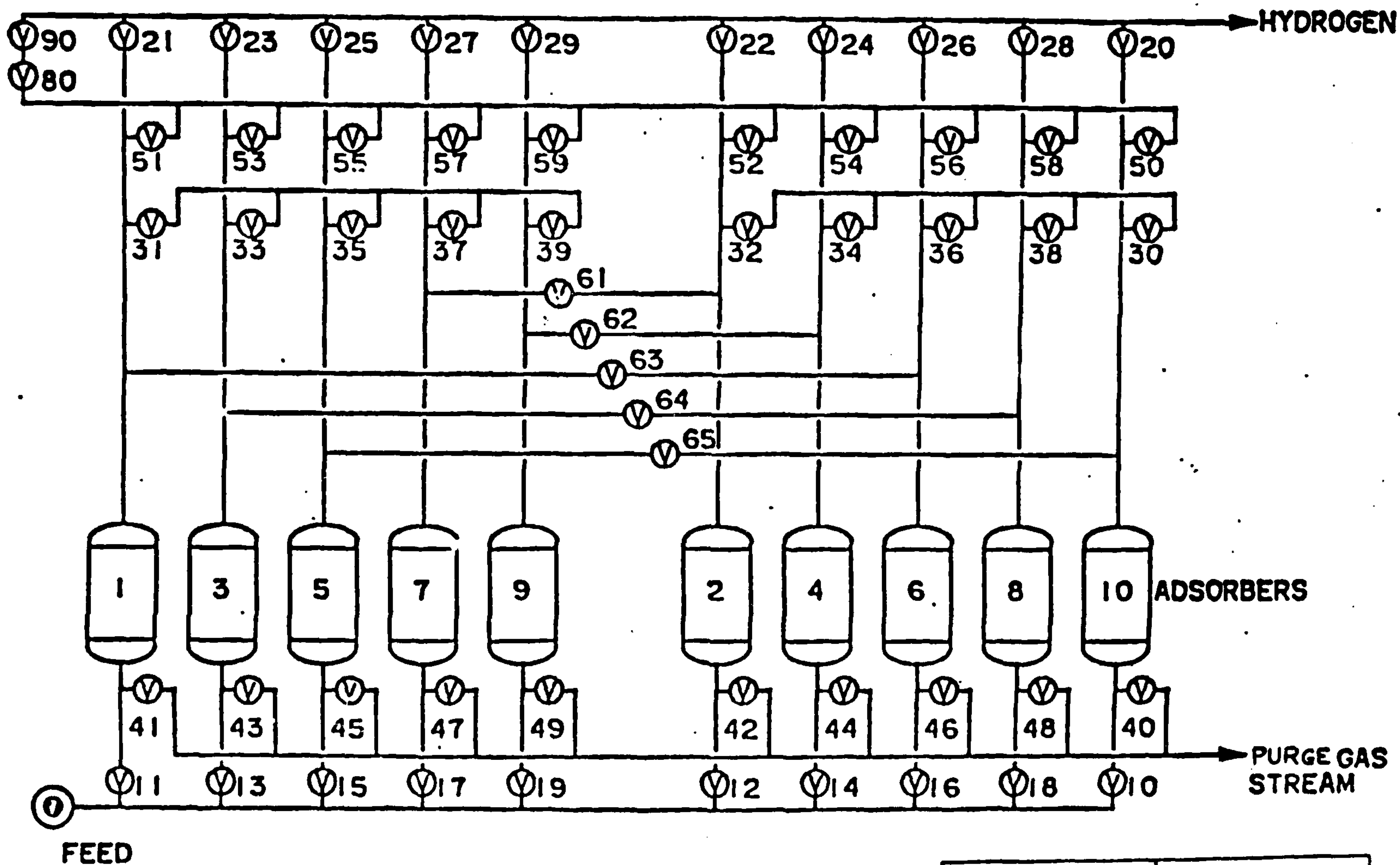
2.0 PROCESS DESCRIPTION

2.2.3 PSA Hydrogen Unit (Area 300) - cont'd

At the conditions given a 4 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

The system uses 4 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-70301 presents a schematic of a Union Carbide Polybed PSA unit.

SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 4

	1/3	FOR SUBTASK 1.2	ML	E.S.		

LUMMUS — THE LUMMUS COMPANY

TITLE PSA HYDROGEN UNIT AREA 300

CLIENT AMOCO/DOE

PROJ NO 5571

CASE 7

5571-70301

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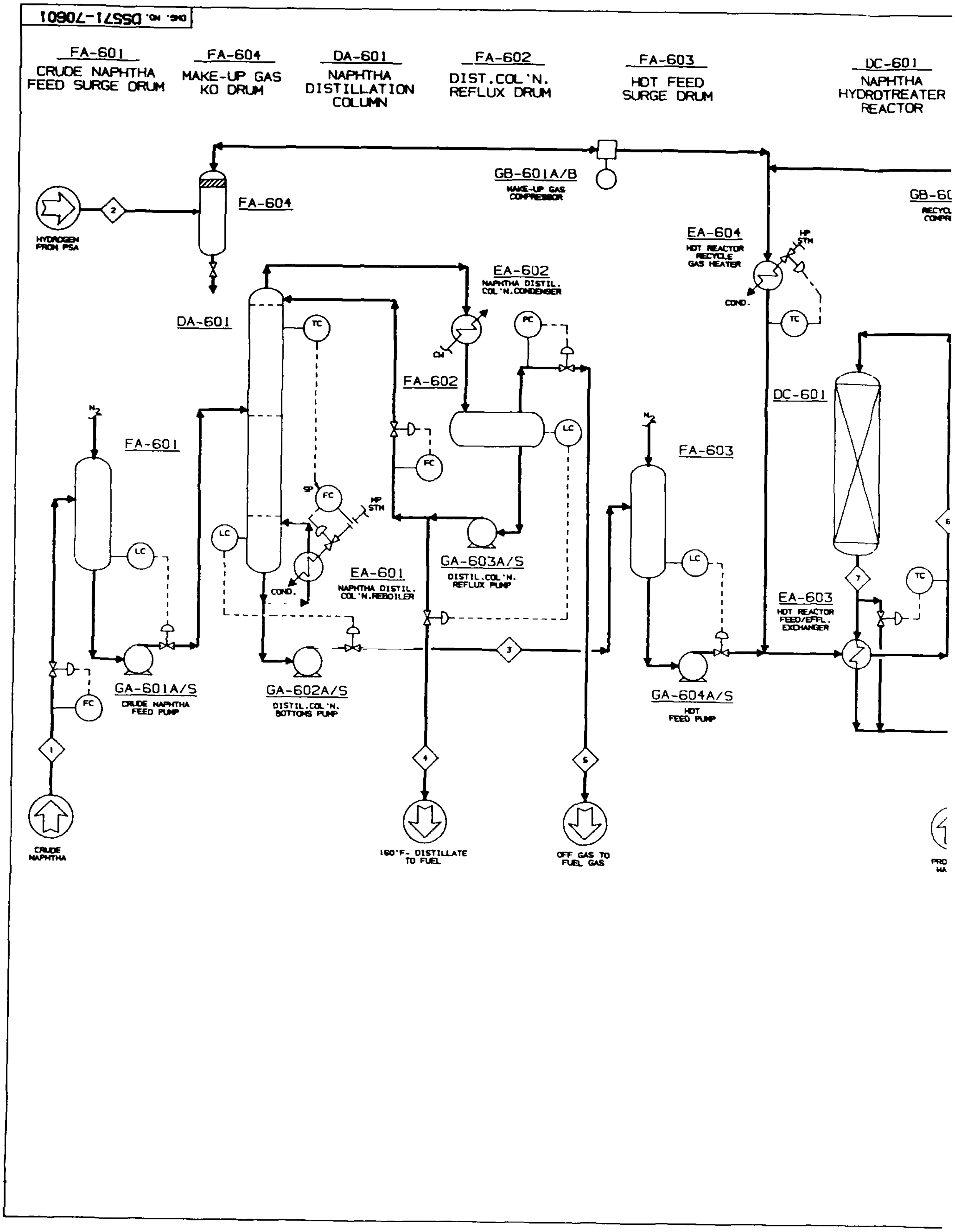
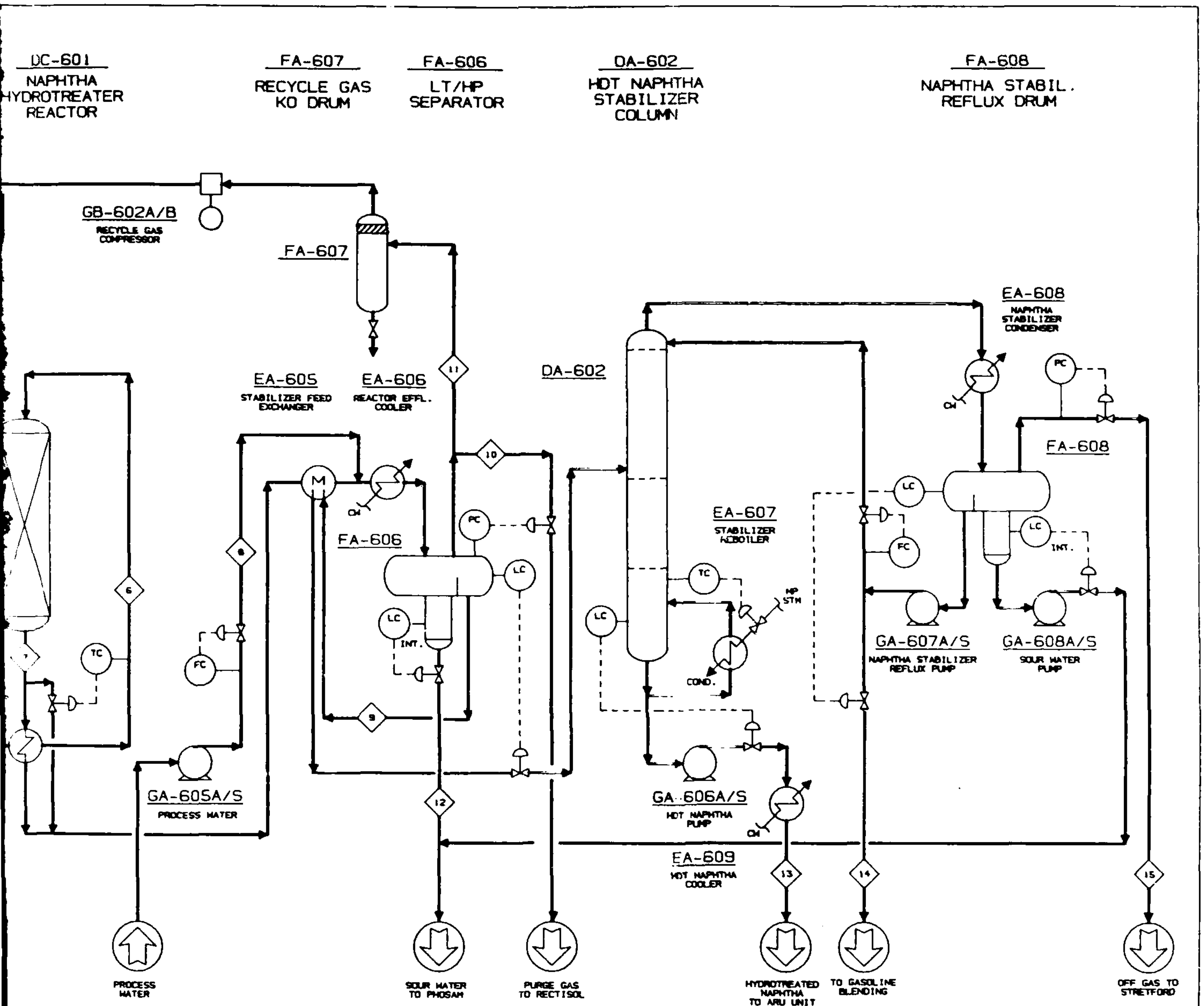

DWG. NO. DSS71-70601
FA-601
CRUDE NAPHTHA FEED SURGE DRUM
FA-604
MAKE-UP GAS KO DRUM
DA-601
NAPHTHA DISTILLATION COLUMN
FA-602
DIST.COL'N. REFLUX DRUM
FA-603
HDT FEED SURGE DRUM
DC-601
NAPHTHA HYDROTREATER REACTOR
GB-601A/B
MAKE-UP GAS COMPRESSOR
HYDROGEN FROM PSA
EA-604
HDT REACTOR RECYCLE GAS HEATER
HP STM
COND.
EA-602
NAPHTHA DISTIL. COL'N.CONDENSER
CW
FA-602
DC-601
N2
FA-601
FA-603
SP
EA-601
NAPHTHA DISTIL. COL'N.REBOILER
GA-603A/S
DISTIL.COL'N. REFLUX PUMP
EA-603
HDT REACTOR FEED/EFFL. EXCHANGER
GA-601A/S
CRUDE NAPHTHA FEED PUMP
GA-602A/S
DISTIL.COL'N. BOTTOMS PUMP
GA-604A/S
HDT FEED PUMP
CRUDE NAPHTHA
160°F- DISTILLATE TO FUEL
OFF GAS TO FUEL GAS



AMOCO/DOE - JET FUELS FROM COAL DERIVED LIQUIDS

COMBUSTION ENGINEERING

LUMMUS CREST INC.
Bloomfield, NJ

PROCESS FLOW DIAGRAM
NAPHTHA DISTILLATION
AND HYDROTREATING
AREA 600 CASE 7

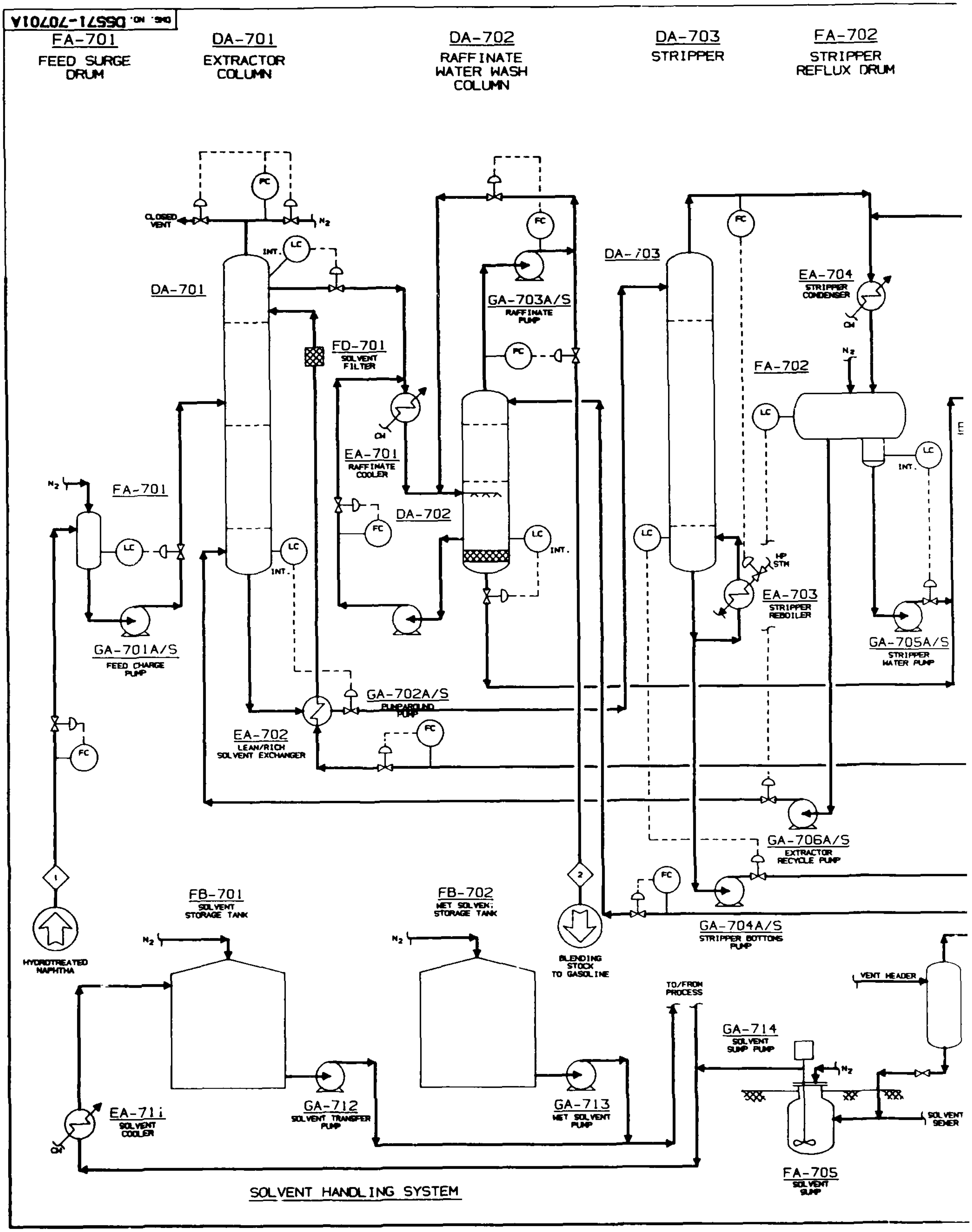
E1466

THIRD ANGLE PROJECTION

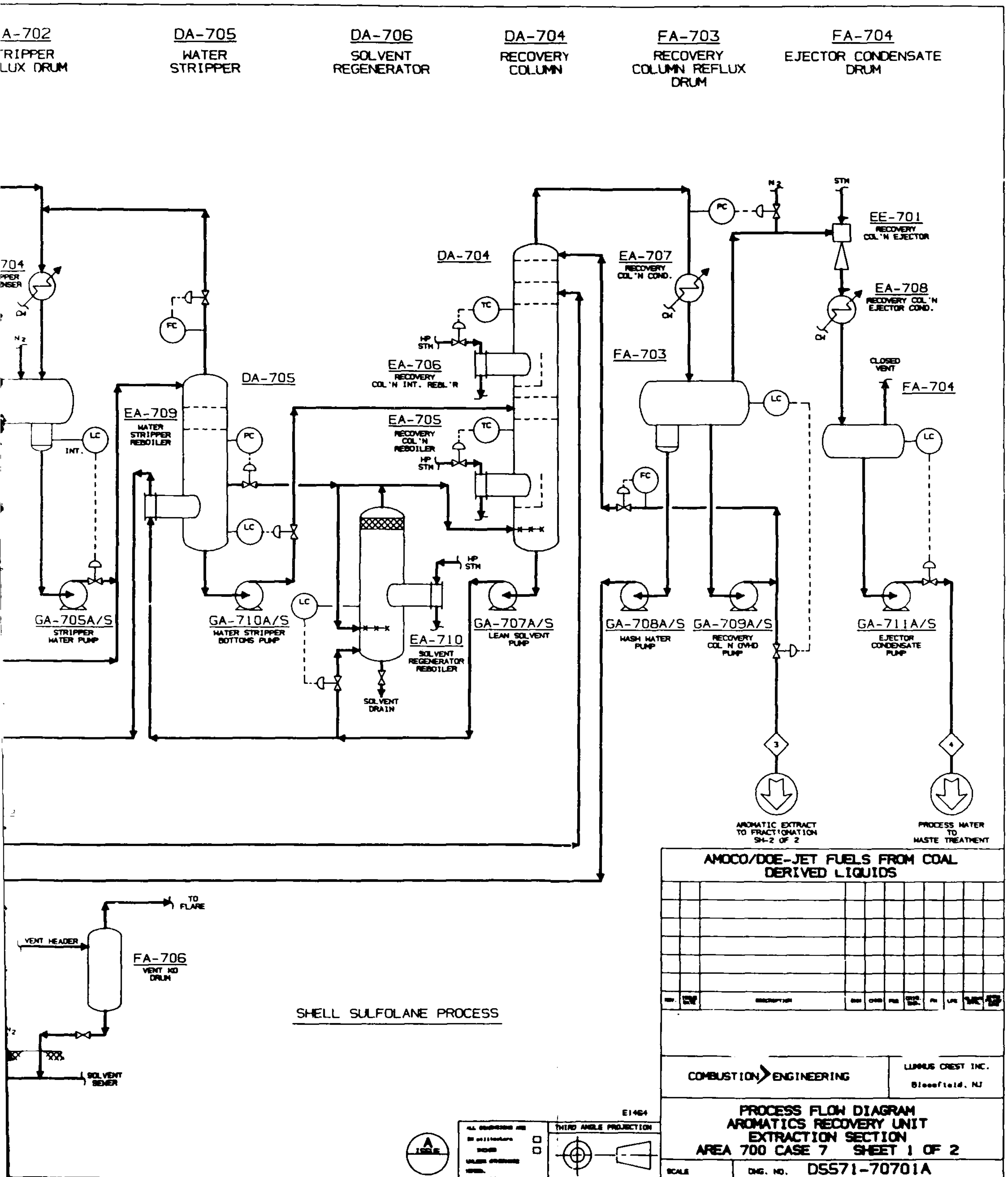
ISSUE A

SCALE

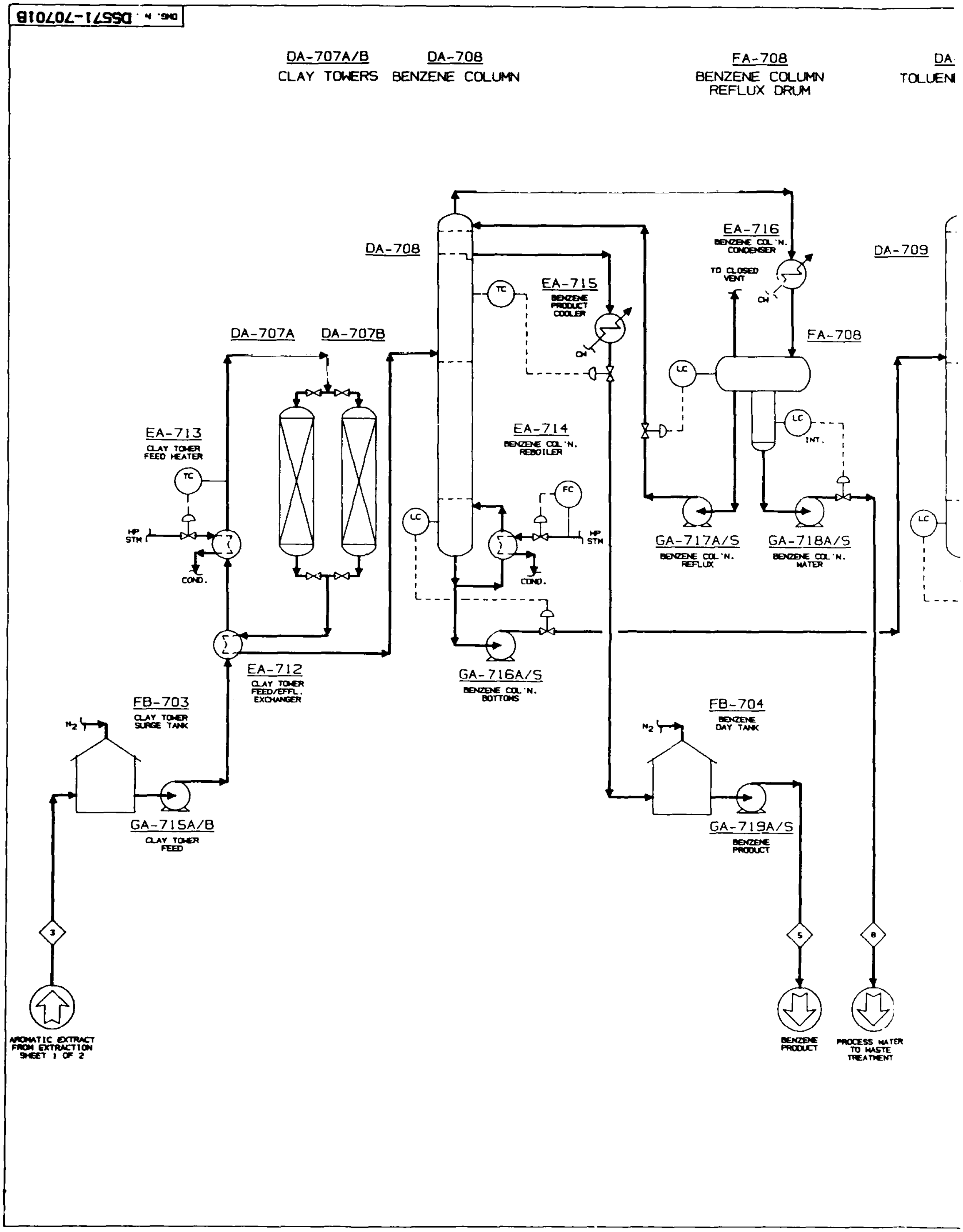
DWG. NO. DS571-70601

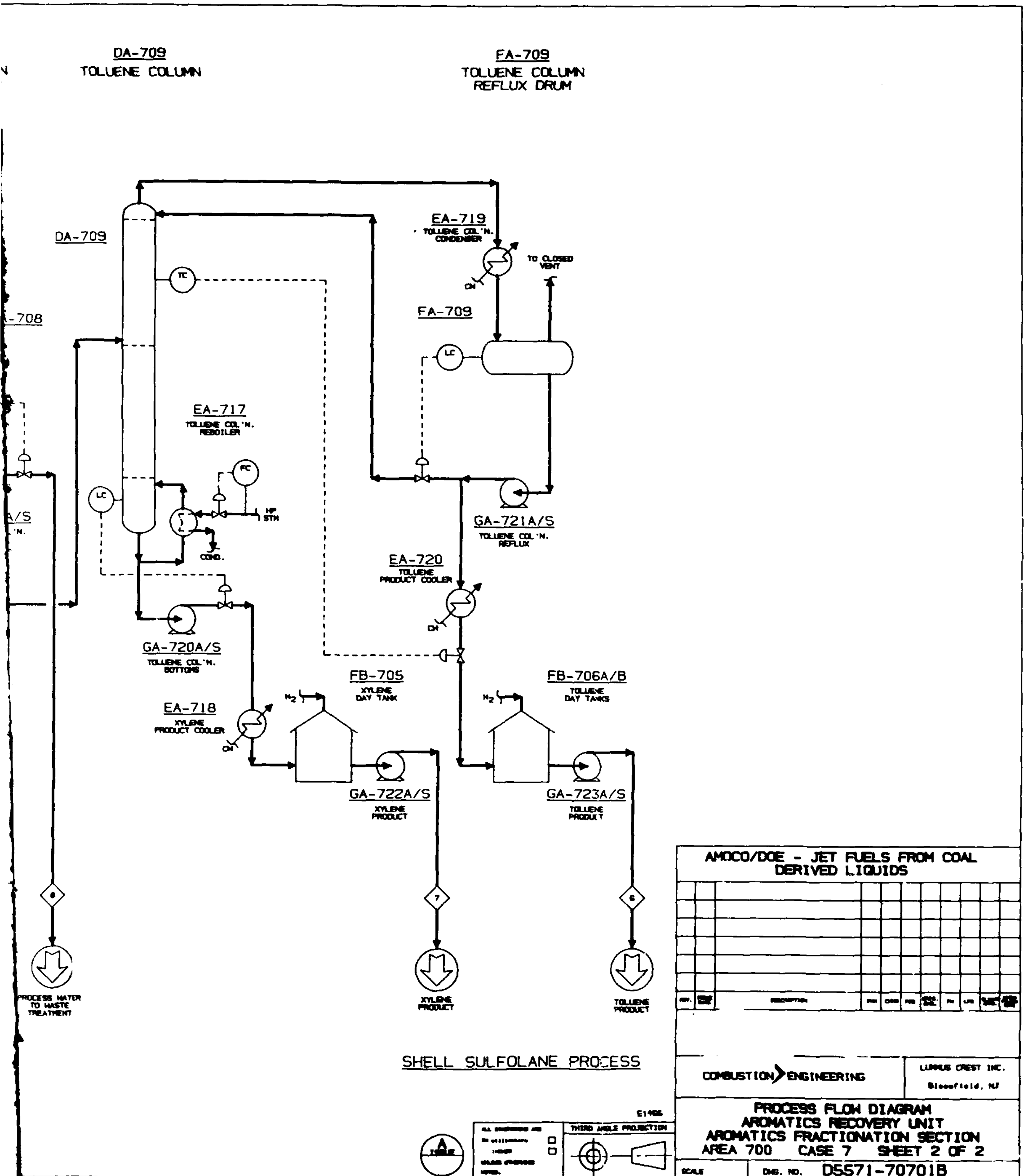


DWG. NO. DSS71-70701A
FA-701
FEED SURGE DRUM
DA-701
EXTRACTOR COLUMN
DA-702
RAFFINATE WATER WASH COLUMN
DA-703
STRIPPER
FA-702
STRIPPER REFLUX DRUM
CLOSED VENT
DA-701
FD-701
SOLVENT FILTER
GA-703A/S
RAFFINATE PUMP
DA-703
EA-704
STRIPPER CONDENSER
FA-702
EA-701
RAFFINATE COOLER
FA-701
DA-702
EA-703
STRIPPER REBOILER
GA-701A/S
FEED CHARGE PUMP
GA-702A/S
PUMPAROUND PUMP
EA-702
LEAN/RICH SOLVENT EXCHANGER
GA-705A/S
STRIPPER WATER PUMP
GA-706A/S
EXTRACTOR RECYCLE PUMP
GA-704A/S
STRIPPER BOTTOMS PUMP
HYDROTREATED NAPHTHA
BLENDING STOCK TO GASOLINE
FB-701
SOLVENT STORAGE TANK
FB-702
WET SOLVENT STORAGE TANK
TO/FROM PROCESS
VENT HEADER
GA-714
SOLVENT SUMP PUMP
EA-711
SOLVENT COOLER
GA-712
SOLVENT TRANSFER PUMP
GA-713
WET SOLVENT PUMP
SOLVENT SEWER
FA-705
SOLVENT SUMP
SOLVENT HANDLING SYSTEM

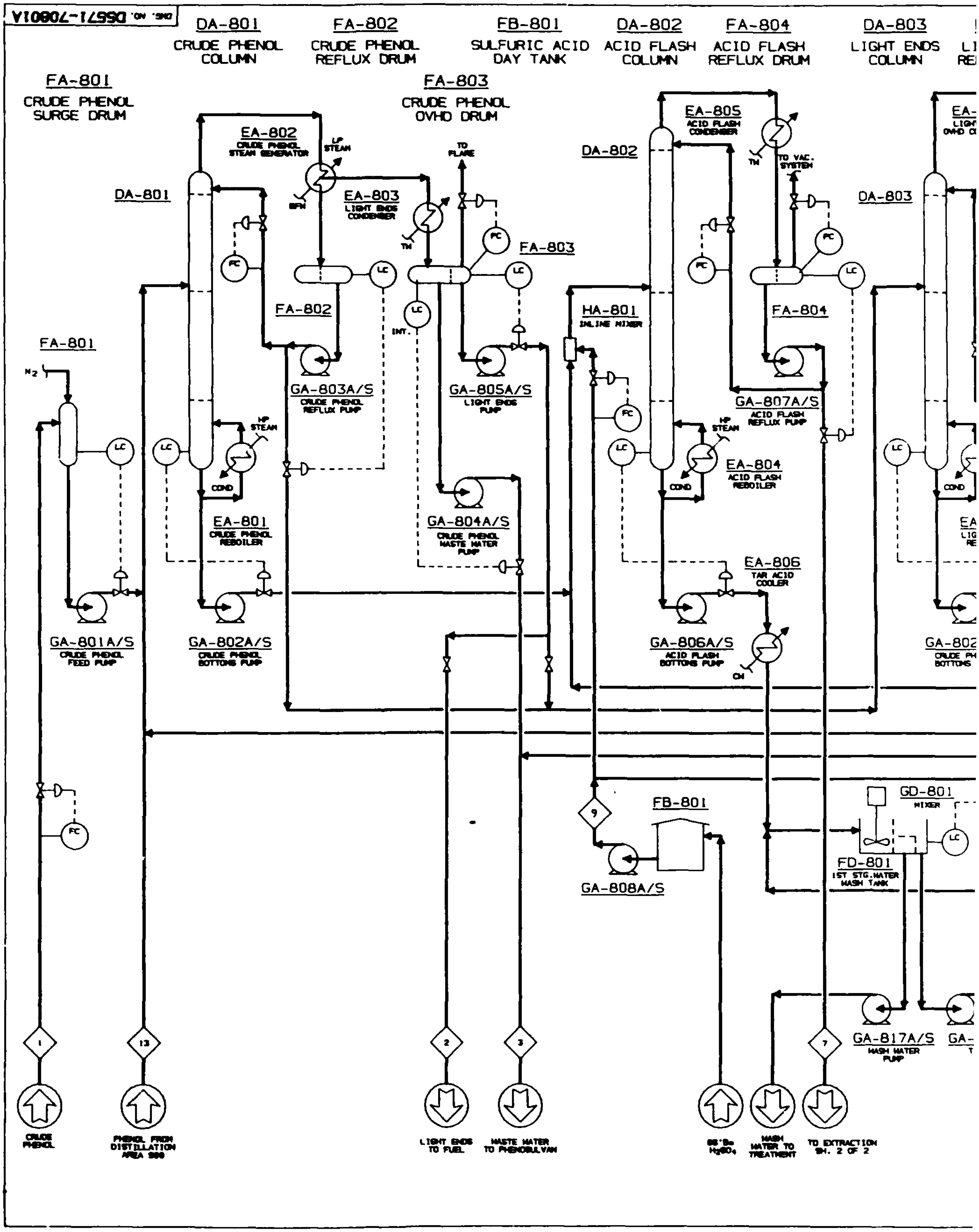


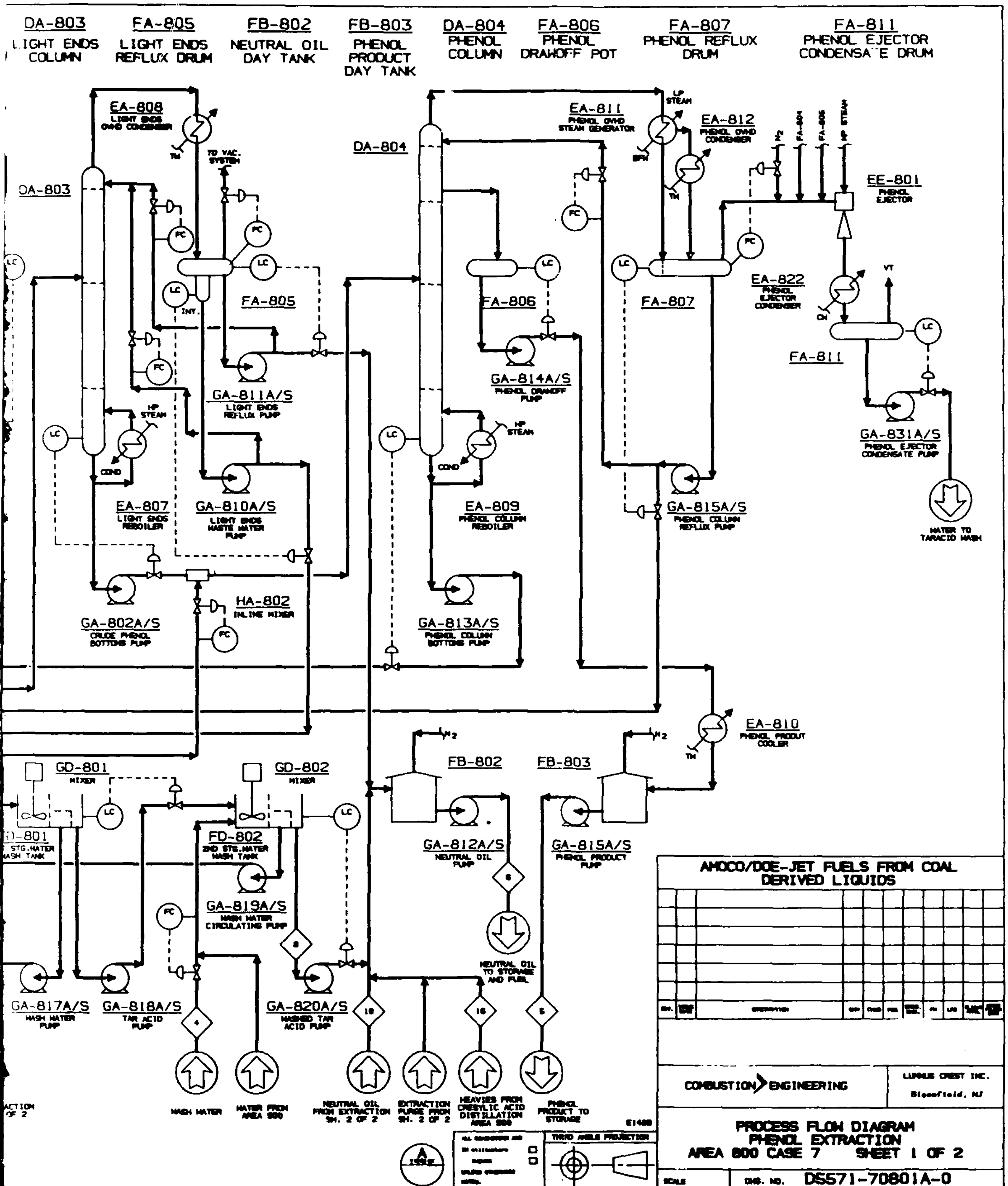
A-702
TRIPPER
LUX DRUM
DA-705
WATER
STRIPPER
DA-706
SOLVENT
REGENERATOR
DA-704
RECOVERY
COLUMN
FA-703
RECOVERY
COLUMN REFLUX
DRUM
FA-704
EJECTOR CONDENSATE
DRUM
EE-701
RECOVERY
COL'N EJECTOR
EA-707
RECOVERY
COL'N COND.
EA-708
RECOVERY COL'N
EJECTOR COND.
DA-704
DA-705
EA-706
RECOVERY
COL'N INT. REBL'R
EA-705
RECOVERY
COL'N
REBOILER
EA-709
WATER
STRIPPER
REBOILER
FA-703
FA-704
CLOSED
VENT
EA-710
SOLVENT
REGENERATOR
REBOILER
SOLVENT
DRAIN
GA-705A/S
STRIPPER
WATER PUMP
GA-710A/S
WATER STRIPPER
BOTTOMS PUMP
GA-707A/S
LEAN SOLVENT
PUMP
GA-708A/S
WASH WATER
PUMP
GA-709A/S
RECOVERY
COL N OVHD
PUMP
GA-711A/S
EJECTOR
CONDENSATE
PUMP
AROMATIC EXTRACT
TO FRACTIONATION
SH-2 OF 2
PROCESS WATER
TO
WASTE TREATMENT
VENT HEADER
TO
FLARE
FA-706
VENT KO
DRUM
SOLVENT
SEWER
SHELL SULFOLANE PROCESS
AMOCO/DOE-JET FUELS FROM COAL
DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
PROCESS FLOW DIAGRAM
AROMATICS RECOVERY UNIT
EXTRACTION SECTION
AREA 700 CASE 7 SHEET 1 OF 2
THIRD ANGLE PROJECTION
E1464
DWG. NO. DS571-70701A


DWG. N. DSS71-70701B
DA-707A/B
CLAY TOWERS
DA-708
BENZENE COLUMN
FA-708
BENZENE COLUMN REFLUX DRUM
DA-
TOLUENE
DA-708
EA-716
BENZENE COL'N. CONDENSER
TO CLOSED VENT
CW
DA-709
TC
EA-715
BENZENE PRODUCT COOLER
CW
FA-708
DA-707A
DA-707B
LC
EA-713
CLAY TOWER FEED HEATER
TC
EA-714
BENZENE COL'N. REBOILER
LC
INT.
FC
LC
HP STM
COND.
HP STM
COND.
GA-717A/S
BENZENE COL'N. REFLUX
GA-718A/S
BENZENE COL'N. WATER
LC
EA-712
CLAY TOWER FEED/EFFL. EXCHANGER
GA-716A/S
BENZENE COL'N. BOTTOMS
FB-703
CLAY TOWER SURGE TANK
N_2
FB-704
BENZENE DAY TANK
N_2
GA-715A/B
CLAY TOWER FEED
GA-719A/S
BENZENE PRODUCT
3
5
8
AROMATIC EXTRACT FROM EXTRACTION SHEET 1 OF 2
BENZENE PRODUCT
PROCESS WATER TO WASTE TREATMENT

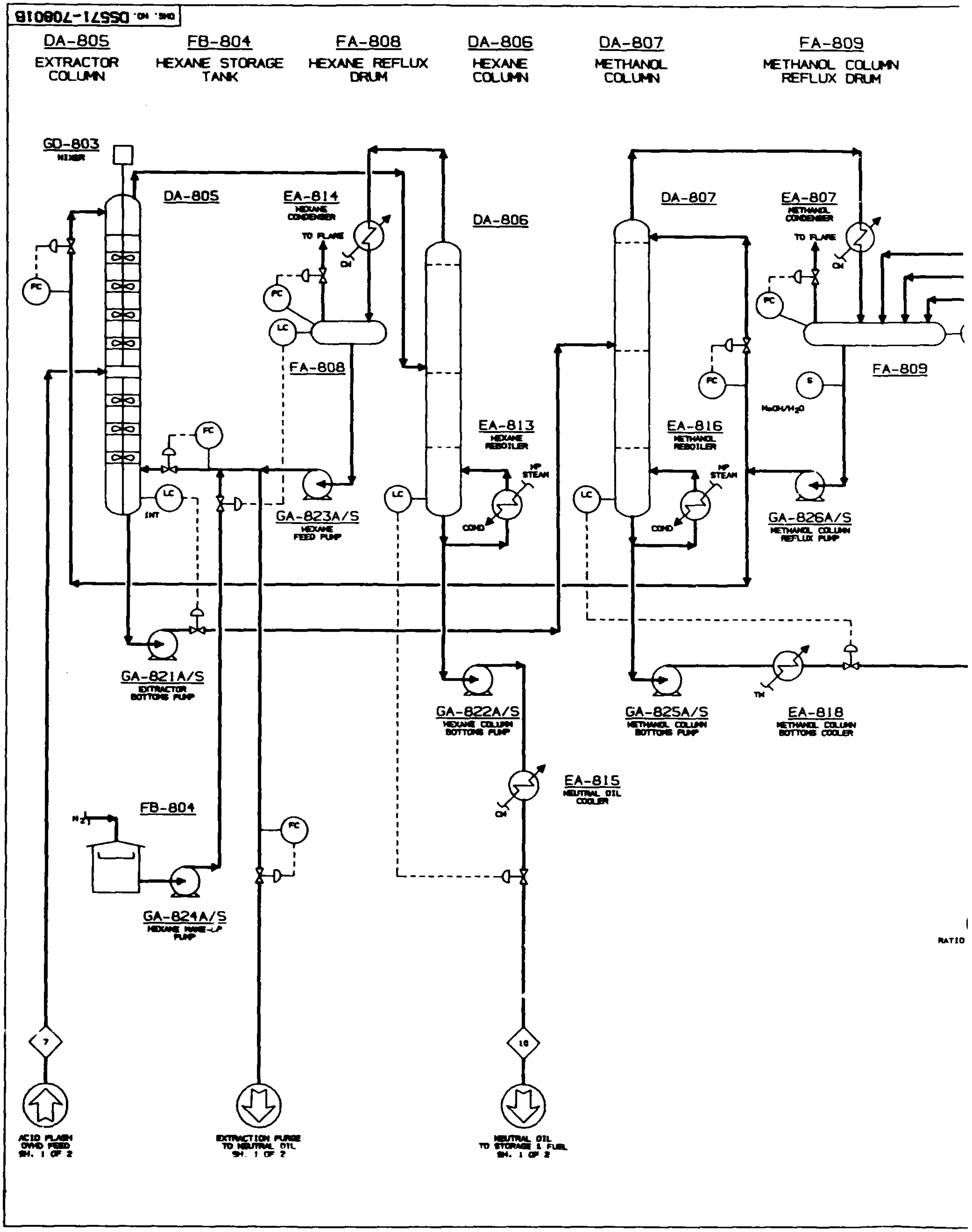


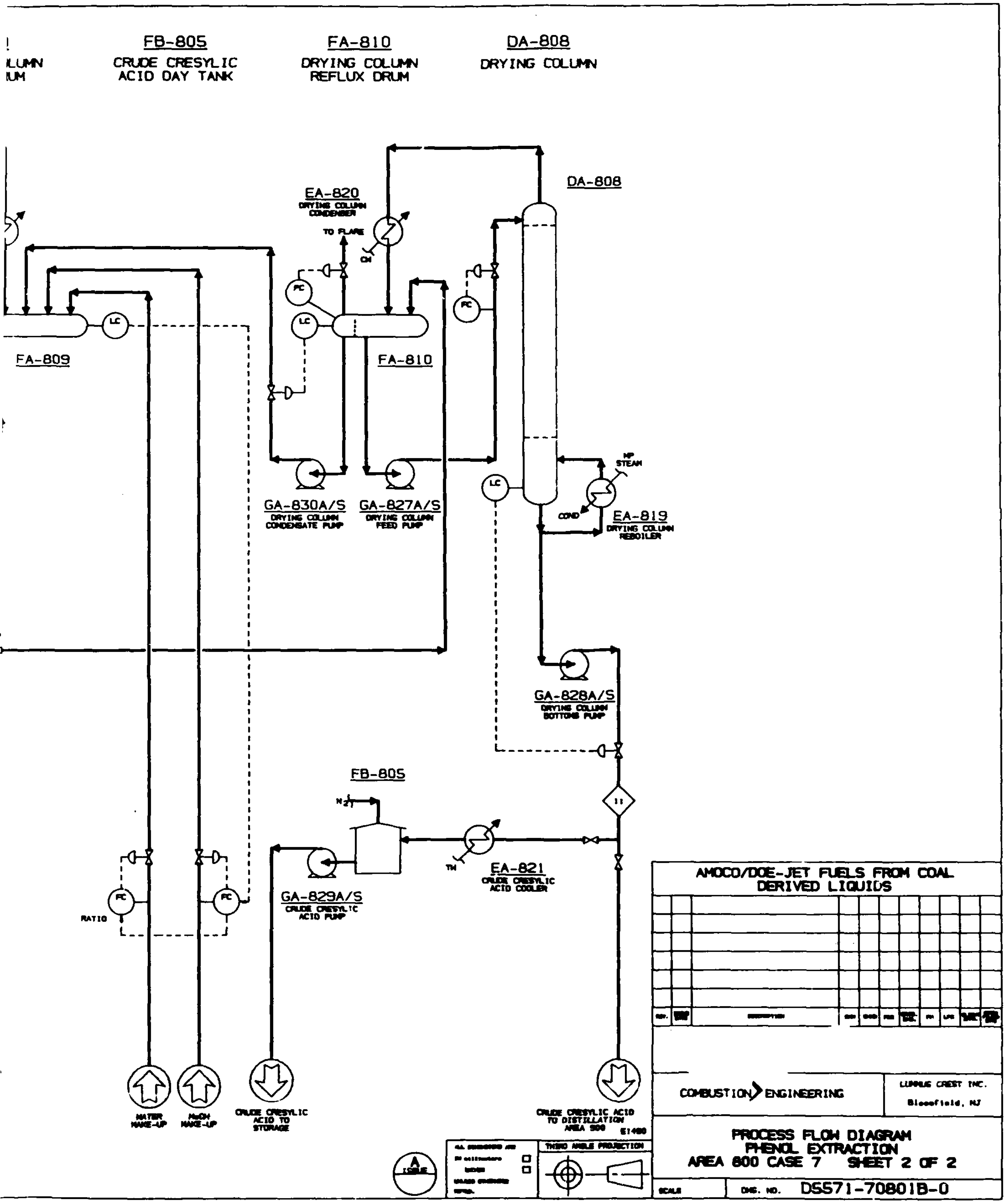
DA-709
TOLUENE COLUMN
FA-709
TOLUENE COLUMN
REFLUX DRUM
DA-709
EA-719
TOLUENE COL'N.
CONDENSER
TO CLOSED
VENT
CW
TC
FA-709
LC
-708
EA-717
TOLUENE COL'N.
REBOILER
FC
LC
HP
STM
COND.
GA-721A/S
TOLUENE COL'N.
REFLUX
EA-720
TOLUENE
PRODUCT COOLER
CW
GA-720A/S
TOLUENE COL'N.
BOTTOMS
FB-705
XYLENE
DAY TANK
FB-706A/B
TOLUENE
DAY TANKS
N_2
N_2
EA-718
XYLENE
PRODUCT COOLER
CW
GA-722A/S
XYLENE
PRODUCT
GA-723A/S
TOLUENE
PRODUCT
8
7
6
PROCESS WATER
TO WASTE
TREATMENT
XYLENE
PRODUCT
TOLUENE
PRODUCT
SHELL SULFOLANE PROCESS
AMOCO/DOE - JET FUELS FROM COAL
DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
PROCESS FLOW DIAGRAM
AROMATICS RECOVERY UNIT
AROMATICS FRACTIONATION SECTION
AREA 700 CASE 7 SHEET 2 OF 2
THIRD ANGLE PROJECTION
SCALE
DWG. NO. D5571-70701B


DWG. NO. DS571-7080IA
DA-801
CRUDE PHENOL COLUMN
FA-802
CRUDE PHENOL REFLUX DRUM
FB-801
SULFURIC ACID DAY TANK
DA-802
ACID FLASH COLUMN
FA-804
ACID FLASH REFLUX DRUM
DA-803
LIGHT ENDS COLUMN
FA-801
CRUDE PHENOL SURGE DRUM
FA-803
CRUDE PHENOL OVHD DRUM
EA-802
CRUDE PHENOL STEAM GENERATOR
EA-803
LIGHT ENDS CONDENSER
EA-805
ACID FLASH CONDENSER
HA-801
INLINE MIXER
GA-803A/S
CRUDE PHENOL REFLUX PUMP
GA-805A/S
LIGHT ENDS PUMP
GA-807A/S
ACID FLASH REFLUX PUMP
EA-804
ACID FLASH REBOILER
EA-801
CRUDE PHENOL REBOILER
GA-804A/S
CRUDE PHENOL WASTE WATER PUMP
EA-806
TAR ACID COOLER
GA-801A/S
CRUDE PHENOL FEED PUMP
GA-802A/S
CRUDE PHENOL BOTTOMS PUMP
GA-806A/S
ACID FLASH BOTTOMS PUMP
GD-801
MIXER
FD-801
1ST STG. WATER WASH TANK
GA-808A/S
GA-817A/S
WASH WATER PUMP
TO FLARE
TO VAC. SYSTEM
LP STEAM
HP STEAM
CRUDE PHENOL
PHENOL FROM DISTILLATION AREA 900
LIGHT ENDS TO FUEL
WASTE WATER TO PHENOSOLVAN
66°Be H2SO4
WASH WATER TO TREATMENT
TO EXTRACTION SH. 2 OF 2

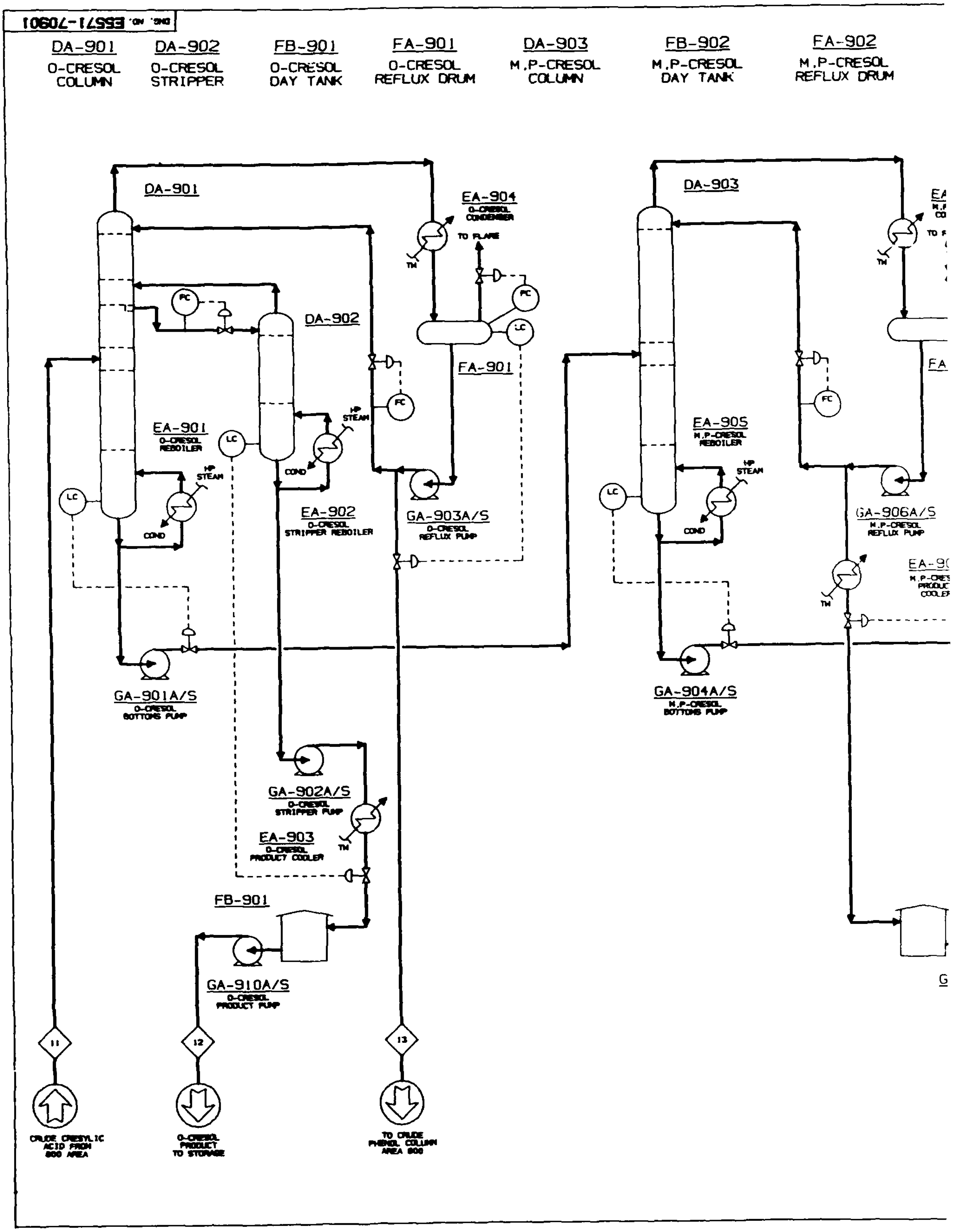


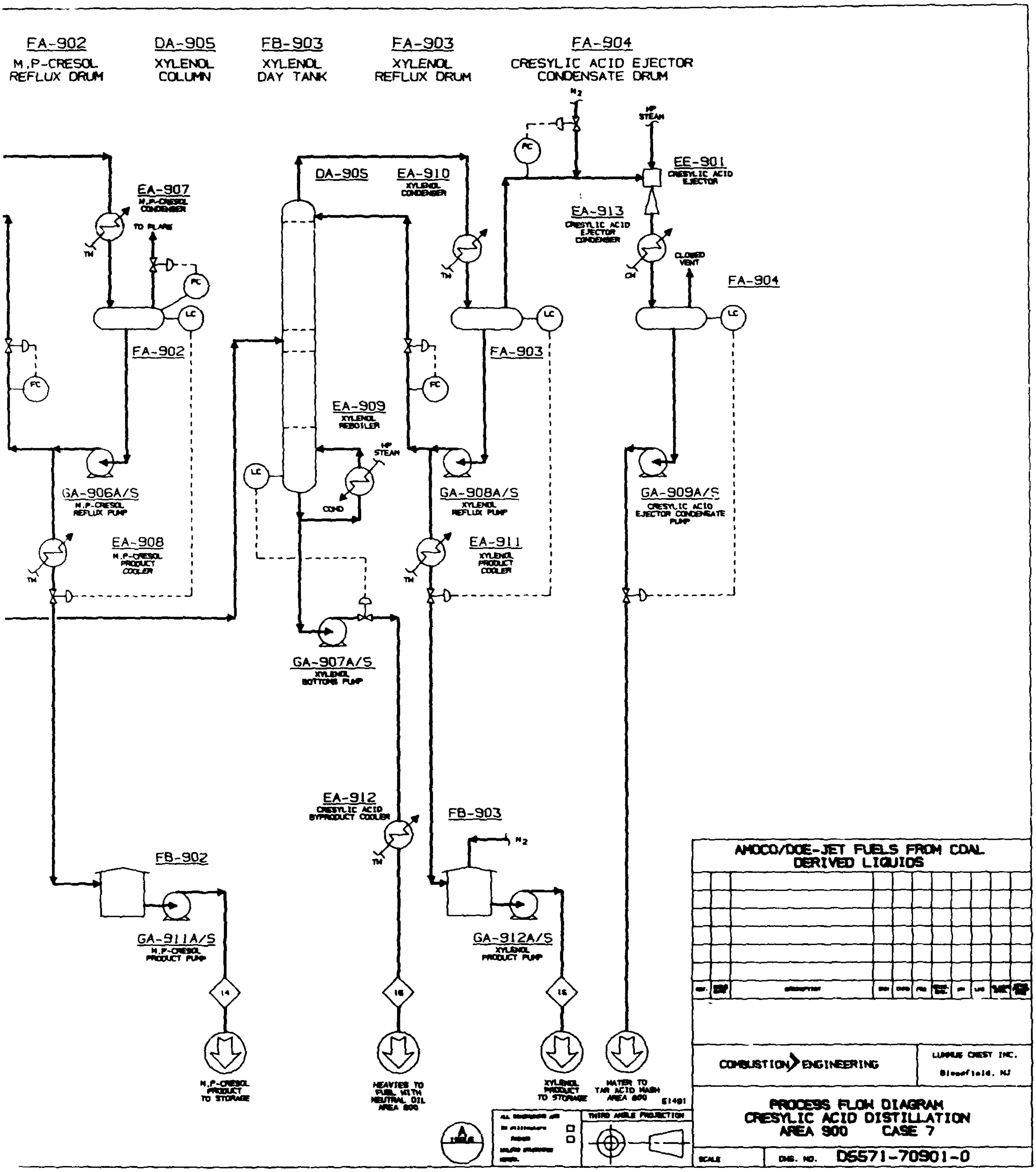
DA-803
LIGHT ENDS COLUMN
FA-805
LIGHT ENDS REFLUX DRUM
FB-802
NEUTRAL OIL DAY TANK
FB-803
PHENOL PRODUCT DAY TANK
DA-804
PHENOL COLUMN
FA-806
PHENOL DRAWOFF POT
FA-807
PHENOL REFLUX DRUM
FA-811
PHENOL EJECTOR CONDENSATE DRUM
EA-808
LIGHT ENDS OVHD CONDENSER
TO VAC. SYSTEM
EA-811
PHENOL OVHD STEAM GENERATOR
EA-812
PHENOL OVHD CONDENSER
LP STEAM
EE-801
PHENOL EJECTOR
EA-822
PHENOL EJECTOR CONDENSER
FA-805
FA-806
FA-807
FA-811
GA-811A/S
LIGHT ENDS REFLUX PUMP
GA-814A/S
PHENOL DRAWOFF PUMP
GA-831A/S
PHENOL EJECTOR CONDENSATE PUMP
WATER TO TARACID WASH
HP STEAM
COND
EA-807
LIGHT ENDS REBOILER
GA-810A/S
LIGHT ENDS WASTE WATER PUMP
EA-809
PHENOL COLUMN REBOILER
GA-815A/S
PHENOL COLUMN REFLUX PUMP
GA-802A/S
CRUDE PHENOL BOTTOMS PUMP
HA-802
INLINE MIXER
GA-813A/S
PHENOL COLUMN BOTTOMS PUMP
EA-810
PHENOL PRODUCT COOLER
GD-801
MIXER
GD-802
MIXER
FB-802
FB-803
FD-802
2ND STG. WATER WASH TANK
GA-812A/S
NEUTRAL OIL PUMP
GA-815A/S
PHENOL PRODUCT PUMP
GA-819A/S
WASH WATER CIRCULATING PUMP
NEUTRAL OIL TO STORAGE AND FUEL
GA-817A/S
WASH WATER PUMP
GA-818A/S
TAR ACID PUMP
GA-820A/S
WASHED TAR ACID PUMP
WASH WATER
WATER FROM AREA 800
NEUTRAL OIL FROM EXTRACTION SH. 2 OF 2
EXTRACTION PURGE FROM SH. 2 OF 2
HEAVIES FROM CRESYLIC ACID DISTILLATION AREA 800
PHENOL PRODUCT TO STORAGE
AMOCO/DOE-JET FUELS FROM COAL DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
AREA 800 CASE 7
SHEET 1 OF 2
THIRD ANGLE PROJECTION
SCALE
DWG. NO. DS571-70801A-0


DA-805
EXTRACTOR COLUMN
FB-804
HEXANE STORAGE TANK
FA-808
HEXANE REFLUX DRUM
DA-806
HEXANE COLUMN
DA-807
METHANOL COLUMN
FA-809
METHANOL COLUMN REFLUX DRUM
GD-803
MIXER
EA-814
HEXANE CONDENSER
TO FLARE
CW
FC
LC
DA-806
EA-807
METHANOL CONDENSER
EA-813
HEXANE REBOILER
HP STEAM
COND
EA-816
METHANOL REBOILER
INT
GA-823A/S
HEXANE FEED PUMP
GA-826A/S
METHANOL COLUMN REFLUX PUMP
GA-821A/S
EXTRACTOR BOTTOMS PUMP
GA-822A/S
HEXANE COLUMN BOTTOMS PUMP
GA-825A/S
METHANOL COLUMN BOTTOMS PUMP
EA-818
METHANOL COLUMN BOTTOMS COOLER
EA-815
NEUTRAL OIL COOLER
FB-804
GA-824A/S
HEXANE MAKE-UP PUMP
RATIO
7
10
ACID FLASH OVHD FEED SH. 1 OF 2
EXTRACTION PURGE TO NEUTRAL OIL SH. 1 OF 2
NEUTRAL OIL TO STORAGE & FUEL SH. 1 OF 2



FB-805
CRUDE CRESYLIC ACID DAY TANK
FA-810
DRYING COLUMN REFLUX DRUM
DA-808
DRYING COLUMN
EA-820
DRYING COLUMN CONDENSER
TO FLARE
CW
DA-808
FC
LC
FA-809
FA-810
FC
LC
HP STEAM
COND
EA-819
DRYING COLUMN REBOILER
GA-830A/S
DRYING COLUMN CONDENSATE PUMP
GA-827A/S
DRYING COLUMN FEED PUMP
GA-828A/S
DRYING COLUMN BOTTOMS PUMP
FB-805
N2
EA-821
CRUDE CRESYLIC ACID COOLER
TW
GA-829A/S
CRUDE CRESYLIC ACID PUMP
FC
FC
RATIO
WATER MAKE-UP
MeOH MAKE-UP
CRUDE CRESYLIC ACID TO STORAGE
CRUDE CRESYLIC ACID TO DISTILLATION AREA 900
E1400
THIRD ANGLE PROJECTION
AMOCO/DOE-JET FUELS FROM COAL DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
AREA 800 CASE 7 SHEET 2 OF 2
SCALE
DWG. NO. DS571-70801B-0


DA-901
O-CRESOL COLUMN
DA-902
O-CRESOL STRIPPER
FB-901
O-CRESOL DAY TANK
FA-901
O-CRESOL REFLUX DRUM
DA-903
M,P-CRESOL COLUMN
FB-902
M,P-CRESOL DAY TANK
FA-902
M,P-CRESOL REFLUX DRUM
DA-901
EA-904
O-CRESOL CONDENSER
TO FLARE
DA-902
FA-901
EA-901
O-CRESOL REBOILER
HP STEAM
COND
EA-902
O-CRESOL STRIPPER REBOILER
GA-903A/S
O-CRESOL REFLUX PUMP
GA-901A/S
O-CRESOL BOTTOMS PUMP
GA-902A/S
O-CRESOL STRIPPER PUMP
EA-903
O-CRESOL PRODUCT COOLER
FB-901
GA-910A/S
O-CRESOL PRODUCT PUMP
CRUDE CRESYLIC ACID FROM 800 AREA
O-CRESOL PRODUCT TO STORAGE
TO CRUDE PHENOL COLUMN AREA 800
DA-903
EA-905
M,P-CRESOL REBOILER
GA-904A/S
M,P-CRESOL BOTTOMS PUMP
GA-906A/S
M,P-CRESOL REFLUX PUMP
M,P-CRESOL PRODUCT COOLER


FA-902
M.P-CRESOL REFLUX DRUM
DA-905
XYLENOL COLUMN
FB-903
XYLENOL DAY TANK
FA-903
XYLENOL REFLUX DRUM
FA-904
CRESYLIC ACID EJECTOR CONDENSATE DRUM
N2
HP STEAM
EE-901
CRESYLIC ACID EJECTOR
EA-907
M.P-CRESOL CONDENSER
TO FLARE
DA-905
EA-910
XYLENOL CONDENSER
EA-913
CRESYLIC ACID EJECTOR CONDENSER
CLOSED VENT
FA-904
FA-902
FA-903
EA-909
XYLENOL REBOILER
HP STEAM
COND
GA-906A/S
M.P-CRESOL REFLUX PUMP
GA-908A/S
XYLENOL REFLUX PUMP
GA-909A/S
CRESYLIC ACID EJECTOR CONDENSATE PUMP
EA-908
M.P-CRESOL PRODUCT COOLER
EA-911
XYLENOL PRODUCT COOLER
GA-907A/S
XYLENOL BOTTOMS PUMP
EA-912
CRESYLIC ACID BYPRODUCT COOLER
FB-903
FB-902
GA-911A/S
M.P-CRESOL PRODUCT PUMP
GA-912A/S
XYLENOL PRODUCT PUMP
14
16
15
M.P-CRESOL PRODUCT TO STORAGE
HEAVIES TO FUEL WITH NEUTRAL OIL AREA 800
XYLENOL PRODUCT TO STORAGE
WATER TO TAR ACID WASH AREA 800
AMOCO/DOE-JET FUELS FROM COAL DERIVED LIQUIDS
COMBUSTION ENGINEERING
LUMMUS CREST INC.
Bloomfield, NJ
THIRD ANGLE PROJECTION
PROCESS FLOW DIAGRAM
CRESYLIC ACID DISTILLATION
AREA 900 CASE 7
SCALE
DWG. NO. D5571-70901-0

3.0 CAPITAL COSTS

3.1 Phenol Stream

3.1.1 Equipment List

Case 7 Maximum Profit

Area 400 -	**Storage Area**
FB-410	Neutral Oil Storage
FB-411	Phenol Product Storage
FB-412	Crude Cresylic Acid Storage
FB-413	O-Cresol Storage
FB-414	M, P Cresol Storage
FB-415	Xylenol Storage
GA-402A/S	Crude Phenol Feed Pump
GA-412A/S	Neutral Oil Transfer Pump
GA-413A/S	Crude Cresylic Acid Transfer Pump
GA-414A/S	O-Cresol Transfer Pump
GA-415A/S	M, P Cresol Transfer Pump
GA-416A/S	Xylenol Transfer Pump
Area 800 -	**Phenol Extraction**
DA-801	Crude Phenol Column
DA-802	Acid Flash Column
DA-803	Light Ends Column
DA-804	Phenol Column
DA-805	Extractor Column
DA-806	Hexane Column
DA-807	Methanol Column
DA-808	Drying Column
EA-801	Crude Phenol Reboiler
EA-802	Crude Phenol Steam Generator
EA-803	LT. Ends Condenser
EA-804	Acid Flash Reboiler
EA-805	Acid Flash Condenser
EA-806	Tar Acid Cooler
EA-807	LT. Ends Reboiler
EA-808	LT. Ends OVHD Condenser
EA-809	Phenol Reboiler
EA-810	Phenol Product Cooler
EA-811	Phenol OVHD Stm. Generator
EA-812	Phenol OVHD Condenser
EA-813	Hexane Reboiler
EA-814	Hexane Condenser
EA-815	Neutral Oil Cooler
EA-816	Methanol Reboiler

3.1 Phenol Stream

3.1.1 Equipment List - cont'd

Area - 800	Phenol Extractor
EA-817	Methanol Condenser
EA-818	Methanol Column Bottoms Cooler
EA-819	Drying Column Reboiler
EA-820	Drying Column Condenser
EA-821	Crude Cresylic Acid Cooler
EA-822	Phenol Ejector Condenser
EE-801	Phenol Ejector
FA-801	Crude Phenol Surge Drum
FA-802	Crude Phenol Reflux Drum
FA-803	Crude Phenol OVHD Drum
FA-804	Acid Flash Reflux Drum
FA-805	Light Ends Reflux Drum
FA-806	Phenol Drawoff Pot
FA-807	Phenol Reflux Drum
FA-808	Hexane Reflux Drum
FA-809	Methanol Reflux Drum
FA-810	Drying Column Reflux Drum
FA-811	Phenol Ejector Condensate Drum
FB-801	Sulfuric Acid Day Tank
FB-802	Neutral Oil Day Tank
FB-803	Phenol Product Day Tank
FB-804	Hexane Storage Tank
FB-805	Crude Cresylic Acid Day Tank
FD-801	1st Stg. Water Wash Tank
FD-802	2nd Stg. Water Wash Tank
GA-801A/S	Crude Phenol Feed Pump
GA-802A/S	Crude Phenol Bottoms Pump
GA-803A/S	Crude Phenol Reflux Pump
GA-804A/S	Crude Phenol Waste Water Pump
GA-805A/S	Light Ends Pump
GA-806A/S	Acid Flash Bottoms Pump
GA-807A/S	Acid Flash Reflux Pump
GA-808A/S	Sulfuric Acid Feed Pump
GA-809A/S	Light Ends Bottoms Pump
GA-810A/S	Light Ends Waste Water Pump
GA-811A/S	Light Ends Reflux Pump
GA-812A/S	Neutral Oil Pump
GA-813A/S	Phenol Column Bottoms Pump
GA-814A/S	Phenol Drawoff Pump
GA-815A/S	Phenol Product Pump
GA-816A/S	Phenol Column Reflux Pump

3.1 Phenol Stream

3.1.1 Equipment List - cont'd

Area 800 - Phenol Extractor

GA-817A/S	Wash Water Pump
GA-818A/S	Acid Tar Pump
GA-819A/S	Wash Water Circulating Pump
GA-820A/S	Washed Tar Acid Pump
GA-821A/S	Extractor Bottoms Pump
GA-822A/S	Hexane Column Bottoms Pump
GA-823A/S	Hexane Feed Pump
GA-824A/S	Hexane Make-up Pump
GA-825A/S	Methanol Column Bottoms Pump
GA-826A/S	Methanol Column Reflux Pump
GA-827A/S	Drying Column Feed Pump
GA-828A/S	Drying Column Bottoms Pump
GA-829A/S	Crude Cresylic Acid Pump
GA-830A/S	Drying Column Condensate Pump
GA-831A/S	Phenol Ejector Condensate Pump
GD-801	1st Stg. Water Wash Mixer
GD-902	2nd Stg. Water Wash Mixer
GD-803	Extractor Mixer
HA-801	Inline Mixer
HA-802	Inline Mixer

Area 900 - Cresylic Acid Distillation

DA-901	O-Cresol Column
DA-902	O-Cresol Stripper
DA-903	M,P-Cresol Column
DA-905	Xylenol Column
EA-901	O-Cresol Reboiler
EA-902	O-Cresol Stripper Reboiler
EA-903	O-Cresol Product Cooler
EA-904	O-Cresol Condenser
EA-905	M,P-Cresol Reboiler
EA-907	M,P-Cresol Condenser
EA-908	M,P-Cresol Product Cooler
EA-909	Xylenol Reboiler
EA-910	Xylenol Condenser
EA-911	Xylenol Product Cooler
EA-912	Cresylic Acid By-product Cooler
EA-913	Cresylic Acid Ejector Condenser
EE-901	Cresylic Acid Ejector

3.1 Phenol Stream

3.1.1 Equipment List - cont'd

Area 900 -	Cresylic Area Distillation
FA-901	O-Cresol Reflux Drum
FA-902	M,P-Cresol Reflux Drum
FA-903	Xylenol Reflux Drum
FA-904	Cresylic Acid Ejector Condensate Drum
FB-901	O-Cresol Day Tank
FB-902	M,P-Cresol Day Tank
FB-903	Xylenol Day Tank
GA-901A/S	O-Cresol Bottoms Pump
GA-902A/S	O-Cresol Stripper Pump
GA-903A/S	O-Cresol Reflux Pump
GA-904A/S	M,P-Cresol Bottoms Pump
GA-906A/S	M,P-Cresol Reflux Pump
GA-907A/S	Xylenol Bottoms Pump
GA-908A/S	Xylenol Reflux Pump
GA-909A/S	Cresylic Acid Ejector Cond. Pump
GA-910A/S	O-Cresol Product Pump
GA-911A/S	M,P-Cresol Product Pump
GA-912A/S	Xylenol Product Pump

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3.1 Phenol Stream - cont'd

3.1.2 Cost Estimate

3.1.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for equipment purchased for various projects. The unit pricing is some what conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the process, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary process basis developed for this section of the plant, a 30% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.1.2.2 Estimate Summary

	(Thousands of $) Case 7
Area 800 Phenol Extraction	12276
Area 900 Cresylic Acid Distil.	4832
Area 400 OSBL (Phenol Stream Part)	3016
Total	$20,124

W16235/bf

3.1 Phenol Stream - cont'd

3.1.2.3 Estimate Breakdown

Area 800

Equipment Items	Type	$ Value	% Comm.	$ Comm.
8	Towers	865	50	432
-	Internals	90	-	-
22	Exchangers	375	100	375
11	Vessels	151	120	181
5	Tanks	74	120	89
2	Filters	30	100	30
62	Pumps	353	120	424
Total 110		$2038		$1531

Equipment	2038	
Commodities	1531	
Labor	1122	10% Equip, 60% Comm.
Indirects	1122	
Office	3630	110 pcs. x 600 x $55
Subtotal	$9,443	
Contingency	2833	30%
	$12,276	

Area 900

Equipment Items	Type	$ Value	% Comm.	$ Comm.
4	Towers	219	120	263
-	Internals	68	-	-
12	Exchangers	138	110	152
4	Vessels	42	120	50
3	Tanks	20	120	24
24	Pumps	132	120	158
Total 47		$ 619		$ 647

Equipment	619	
Commodities	647	
Labor	450	10% Equip, 60% Comm.
Indirects	450	
Office	1551	47 x 600 x $55
Subtotal	$3,717	
Contingency	1,115	30%
	$4,832	

W16235/bf

3.1 Phenol Stream - cont'd

3.1.2.3 Estimate Breakdown - cont'd

Area 400

Tankage		MTL	S/C	$199
Pumps		MTL		54
Equipment related commodities				200
Yard Piping	29000 LF	MTL		$330
	Labor			870
Pipe Insul.		S/C		178
Tracing		S/C		132
Excavation		S/C		150
Rack		S/C		300
	Subtotal			$2,413
	Contingency	25%		603
	Total			$3,016

W16235/bf

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL RERIVED LIQUIDS

3.0 CAPITAL COSTS

3.2 Naphtha Stream

3.2.1 Equipment List

Case 7 - Maximum Profit

Area 300 - PSA Hydrogen Unit

Tag. No.	Description
PA-301	PSA Hydrogen Unit Package

Area 400 - Storage

Tag. No.	Description
FB-403	Fuel Oil Storage
FB-404	Blending Stock
FB-405	Benzene Storage
FB-406	Toluene Storage
FB-407	Xylene Storage
FB-408	Butane Storage
FB-409	Gasoline Storage
GA-403A/S	Fuel Oil Transfer Pump
GA-405A/S	Crude Naphtha Pump
GA-406A/S	Blending Stock Pump
GA-407A/S	Benzene Transfer Pump
GA-408A/S	Toluene Transfer Pump
GA-409A/S	Xylene Transfer Pump
GA-410A/S	Butane Transfer Pump
GA-411A/S	Gasoline Transfer Pump
PA-401	Gasoline Blending Package

Area 600 - Naphtha Distillation & HDT

Tag. No.	Description
DA-601	Naphtha Distillation Column
DA-602	HDT Naphtha Stabilizer Column
DC-601	Naphtha Hydrotreater Reactor

W16235/bf

3.2 Naphtha Stream

3.2.1 Equipment List - cont'd

Area 600	- Naphtha Distillation & HDT
EA-601	Naphtha Distillation Column Reboiler
EA-602	Naphtha Distillation Column Condenser
EA-603	HDT Reactor Feed/Effl. Exchanger
EA-604	HDT Reactor Recycle Gas Heater
EA-605	Stabilizer Feed Exchanger
EA-606	Reactor Effl. Cooler
EA-607	Stabilizer Reboiler
EA-608	Naphtha Stabilizer Condenser
EA-609	HDT Naphtha Cooler
FA-601	Crude Naphtha Feed Surge Drum
FA-602	Distil. Col'n Reflux Drum
FA-603	HDT Feed Surge Drum
FA-604	Makeup Gas KO Drum
FA-606	LT/HP Separator
FA-607	Recycle Gas KO Drum
FA-608	Naphtha Stabil. Reflux Drum
GA-601A/S	Crude Naphtha Feed Pump
GA-602A/S	Distil. Col'n Bottoms Pump
GA-603A/S	Distil. Col'n Reflux Pump
GA-604A/S	HDT Feed Pump
GA-605A/S	Process Water Pump
GA-606A/S	HDT Naphtha Pump
GA-607A/S	Naphtha Stabil. Reflux Pump
GA-608A/S	Sour Water Pump
GB-601A/B	Makeup Gas Compressor
GB-602A/B	Recycle Gas Compressor

3.2 Naphtha Stream

3.2.1 Equipment List - cont'd

Area 700 -	Aromatics Recovery
DA-701	Extractor Column
DA-702	Raffinate Water Wash Column
DA-703	Stripper
DA-704	Recovery Column
DA-705	Water Stripper
DA-706	Solvent Regenerator
DA-707A/B	Clay Tower
DA-708	Benzene Column
DA-709	Toluene Column
EA-701	Raffinate Cooler
EA-702	Lean/Rich Solvent Exchanger
EA-703	Stripper Reboiler
EA-704	Stripper Condenser
EA-705	Recovery Column Reboiler
EA-706	Recovery Column Intermediate Reboiler
EA-707	Recovery Column Condenser
EA-708	Recovery Column Ejector Condenser
EA-709	Water Stripper Reboiler
EA-710	Solvent Regenerator Reboiler
EA-711	Solvent Cooler
EA-712	Clay Tower Feed/Effl. Exchanger
EA-713	Clay Tower Feed Heater
EA-714	Benzene Column Reboiler
EA-715	Benzene Product Cooler
EA-716	Benzene Column Condenser
EA-717	Toluene Column Reboiler
EA-718	Xylene Product Cooler
EA-719	Toluene Column Condenser
EA-720	Toluene Product Cooler
EE-701	Recovery Column Ejector
FA-701	Feed Surge Drum
FA-702	Stripper Reflux Drum
FA-703	Recovery Column Reflux Drum
FA-704	Ejector Condensate Drum
FA-705	Solvent Sump
FA-706	Vent KO Drum
FA-708	Benzene Column Reflux Drum
FA-709	Toluene Column Reflux Drum

W16235/bf

3.2 Naphtha Stream

3.2.1 Equipment List - cont'd

FB-701	Solvent Storage Tank
FB-702	Wet Solvent Storage Tank
FB-703	Clay Tower Surge Tank
FB-704	Benzene Day Tank
FB-705	Xylene Day Tank
FB-706A/B	Toluene Day Tanks
FD-701	Solvent Filter
GA-701A/S	Feed Charge Pump
GA-702A/S	Pumparound Pump
GA-703A/S	Raffinate Pump
GA-704A/S	Stripper Bottoms Pump
GA-705A/S	Stripper Water Pump
GA-706A/S	Extractor Recycle Pump
GA-707A/S	Lean Solvent Pump
GA-708A/S	Wash Water Pump
GA-709A/S	Recovery Column OV'HD Pump
GA-710A/S	Water Stripper Bottoms Pump
GA-711A/S	Ejector Condensate Pump
GA-712	Solvent Transfer Pump
GA-713	Wet Solvent Pump
GA-714A/B	Solvent Sump Pump (Warehouse Spare)
GA-715A/S	Clay Tower Feed Pump
GA-716A/S	Benzene Column Bottoms Pump
GA-717A/S	Benzene Column Reflux Pump
GA-718A/S	Benzene Column Water Pump
GA-719A/S	Benzene Product Pump
GA-720A/S	Toluene Column Bottoms Pump
GA-721A/S	Toluene Column Reflux Pump
GA-722A/S	Xylene Product Pump
GA-723A/S	Toluene Product Pump
PA-701	Clay Handling Equipment

3.2.2 Cost Estimate

3.2.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for various projects. The unit pricing is some what conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

3.2.2 Cost Estimate

3.2.2.1 Basis of Estimate - cont'd

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the high pressure processing, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2.2 Estimate Summary

(Thousands of $)

	Case 7
Area 600 Naphtha Distill. & HDT	4615
Area 700 Aromatics Recovery	9373
Area 300 PSA	518
Area 400 OSBL (Naphtha Stream Part)	3058
Subtotal	$17,564
Area 700 - ARU Solvent Inventory	100
Total	$17,664

W16235/bf

3.2 Naphtha Stream - cont'd

3.2.2.3 Estimate Breakdown (Area 600) All values in Thousands

	Equipment Items	Type	$ Value	% Comm.	$ Comm.
	2	Towers	48	140	67
	-	Internals	8	-	-
	1	Exchangers	125	85	106
	9	Vessels	123	100	123
	7	Tanks	89	100	89
	16	Pumps	68	100	68
	4	Compressors	230	60	138
Total	39		$ 691		$ 591

Equipment	691	
Commodities	591	
Labor	424	10% Equip., 60% Comm.
Indirects	424	
Office	1716	39 pcs. x 800 x $55
Subtotal	$3,846	
Contingency	769	20%
Total	$4,615	

Area 700

	Equipment Items	Type	$ Value	% Comm.	$ Comm.
	10	Towers	350	140	490
	8	Internals	66	-	-
	20	Exchangers	113	100	113
	9	Vessels	65	120	78
	7	Tanks	117	100	117
	44	Pumps	180	120	216
	3	Special	20	100	20
Total	101		$ 911		$1034

Equipment	911	
Commodities	1034	
Labor	711	10% Equip., 60% Comm.
Indirects	711	100%
Office	4444	101 pcs. x 800 x $55
Subtotal	$7,811	
Contingency	1,562	20%
Total	$9,373	

W16235/bf

3.2 Naphtha Stream - cont'd

3.2.2.3 Estimate Breakdown (Area 600) All Values in Thousands - cont'd

Area 300

PSA Hydrogen Package Unit (one skid)

205 MMSCFD budget quote -		$300
Installation	50%	150
Subtotal		$450
Contingency	15%	68
Total		$518

Area 400

Equipment & Value

Tankage	MTL	S/C	366
Pumps	MTL		62
Yard Piping 27,000 LF	MTL		300
Labor	S/C		813
Pipe Insulation	S/C		108
Tracing	S/C		84
Excavation	S/C		135
Rack	S/C		300
Equipment Related Commodities 65% Equip.			278
	Subtotal		$2,446
	Contingency 25%		612
			$3,058

W16235/bf

4.0 OPERATING COSTS

4.1 Phenol Stream

4.1.1 Operating Labor

It is estimated that it will require 6 men/shift to operate the plant broken down as follows:

Foreman	1
Control Room	1
Phenol Extr. Operator	1
Crsylic Acid Distill. Operator	1
	4 Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 3 people.

The total additional people (assuming 4 & 2 operation for the process units) are as follows:

Shift Personnel	4 positions x 4 people/ position -	16
Supervisor & Admin.		5
QC Technician		1
Maintenance		3
Other (Stores or Janitorial)		1
Total		26

4.1.2 Utilities

The following utilities have been estimated from the preliminary process designs:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	626 BPSD	$16/Bbl [a]	10,016
Cooling Water	2300 GPM	$0.155/MGal [b]	513
Power	290 kW	$0.04/kWH [b]	278
Process Water	2 GPM	$J.45/MGal [b]	2

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) ANG utility cost information dated 5/87.

W16235/bf

4.1.3 Chemicals

The chemical cost is as follows:

Chemical	Use	Cost	$/SD
H_2SO_4	7100 #/D	$0.04/#	285

4.1.4 Maintenance Supplies

Maintenance supplies costs are not known but will be assumed to be 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

0.00005 x 20,124,000 = $1006/SD

4.2 Naphtha Stream

4.2.1 Operating Labor

It is estimated that it will require 7 men/shift to operate this section plant broken down as follows:

Foreman	1	
Control Room	1	
HDT Operator	2	
ARU Operator	2	
PSA & relief man	1	
	7	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 5 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	7 positions x 4 people/ position -	28
Supervisor & Admin.		5
QC Technician		1
Maintenance		5
Other (Stores or Janitorial)		1
Total		40

4.2.2 Utilities

The following utilities have been estimated from the simulations:

Utility	Consumption	Cost	$/SD
#6 Fuel Oil	567 BPSD	$16/Bbl (a)	9072
SNG equivalent	0.17 MM SCFD	$3.80/MM BTU (b)	633
Cooling Water	1700 GPM	$0.155/MGal (c)	380
Power	235 kW	$0.04/kWH (c)	226
Process Water	1 GPM	$0.45/MGal (c)	1

(a) Cut of 1% sulfur $6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) Memo from D. Daley of BRSC to L. Lorenzo of DOE dtd. 10/20/87, Ref. Dpd-87-863.

(c) ANG utility cost information dated 5/87.

4.2.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

Catalyst &Chem	Use	Cost	$/SD
HDT Cat.	0.021 #/Bbl	$3.00/#	33
ARU Solvent	24 #/D	$2.10/#	50
			$88

4.2.4 Maintenance Supplies

Maintenance supplies for refining operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units excluding solvent inventory. On this basis the maintenance supplies would be

$$0.00005 \times 17{,}564{,}000 = \$878/SD$$

W16235/bf

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the maximum profit case are proposed to be located to the east of the Phenosolvan and Waste Water Units of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 375' x 350' will be surrounded by an access road and will be divided by a central east-west road. Areas 300, 600 & 700 will be located to the south and Areas 800 & 900 to the north of the road.

A diked storage tank area approx. 360' x 260' will be required for product and fuel oil storage associated with the naphtha stream and an area approx. 60' x 300' is required for the phenol stream. These areas are proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

Approximately 1200 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines is shown in table 5.1.

W16235/bf

TABLE 5.1

INTERCONNECTING PIPING

I - PHENOL STREAM

EST. SIZE	SERVICE	TO/FROM
2"	Crude Phenol (Elec. Tr.)	Storage
1 1/2"	Light Ends	Storage/Ph. Ext.
1 1/2"	Neutral Oil	Storage
1 1/2"	Phenol Product (Elec. Tr.)	Storage
1 1/2"	Crude Cresylic Acid (Elec. Tr.)	Storage
1 1/2"	O-Cresol (Elec. Tr.)	Storage
1 1/2"	M,P-Cresol (elec. Tr.)	Storage
1 1/2"	Xylenol Product (Elec. Tr.)	Storage
1 1/2"	Methanol Make-up	Ph. Ext./MeoH Unit
1 1/2"	Sulfuric Acid (Elec. Tr.)	Ph. Ext/Storage
12"	Wet Flare (Trace)	Flare
2"	Nitrogen	Main rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
3"	LP Steam	"
2"	M.P. Steam	"
6"	H.P. Steam	"
4"	Stm. Cond.	"
2"	BFW	"
12"	C. W. Supply & Return	"
1 1/2"	Wash Water	Treatment/Ph. Ext.
1 1/2"	Waste Water (Elec. Tr.)	Phenosolvan/Ph. Ext.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers

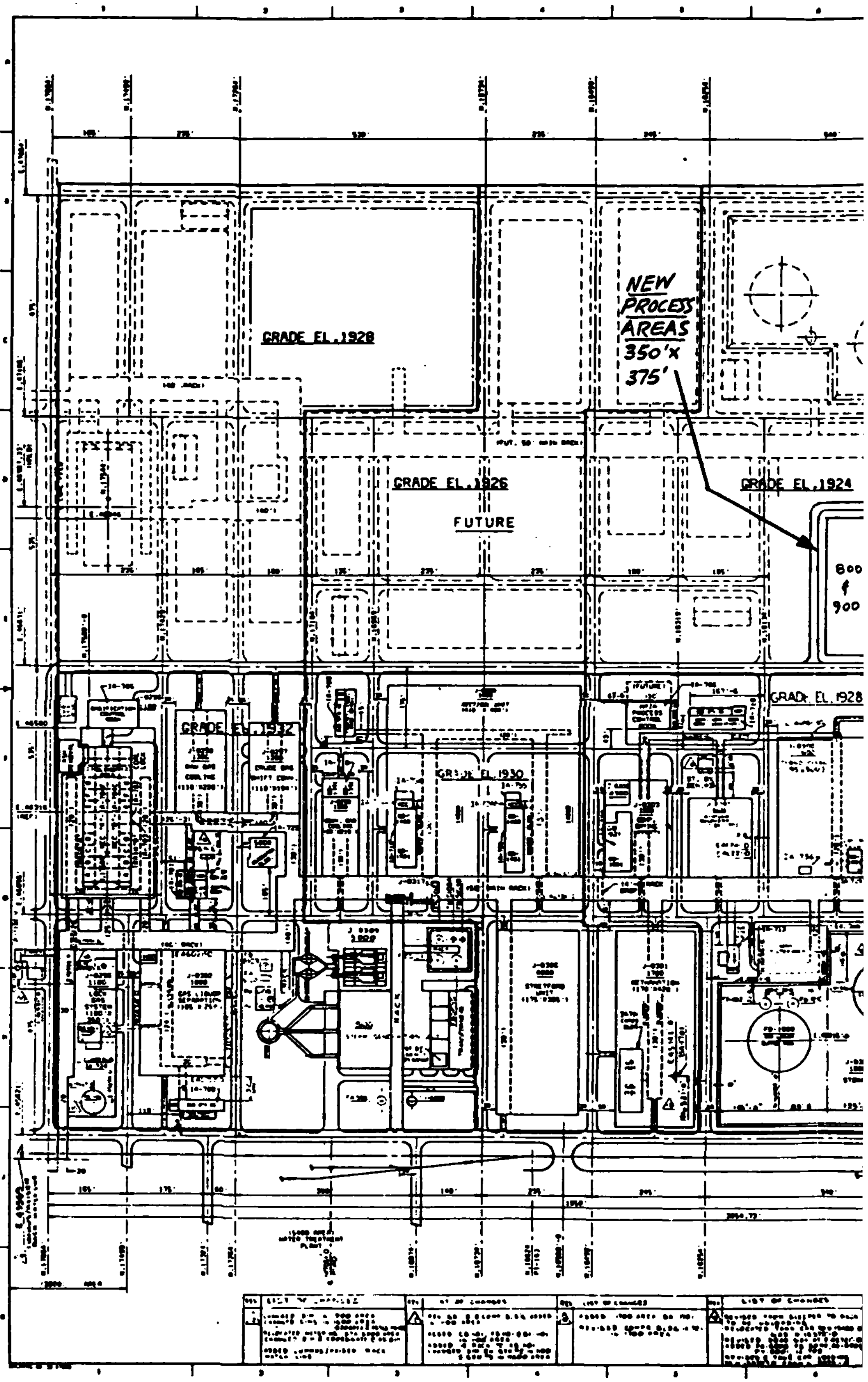
TABLE 5.1 - cont'd

INTERCONNECTING PIPING

II - NAPHTHA STREAM

EST. SIZE	SERVICE	TO/FROM
1 1/2	Crude Naphtha	Storage
1 1/2"	160°F - Distillate	Storage/Dist.
1 1/2"	Blending Stock	Storage/ARU
1 1/2"	Benzene	Storage/ARU
1 1/2"	Toluene	Storage/ARU
1 1/2"	Xylene	Storage/ARU
1 1/2"	Butane	Storage
3"	Gasoline	Storage
14"	Wet Flare (Trace)	Flare
1 1/2"	Synthesis Gas	PSA/Rectisol
3"	Purge Gas	Fuel Gas/PSA & HDT
1 1/2"	Off Gas	Rectisol/HDT
2"	Nitrogen	Main Rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
6"	M.P. Steam	"
6"	H.P. Steam	"
6"	Stm Cond.	"
10"	C. W. Supply & Return	"
2"	Waste Water	Phosam/HDT
4"	Fuel Oil (Elec. Tr.)	Exist TKS/New TKS.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers

W16235/bf


NEW PROCESS AREAS 350'x 375'
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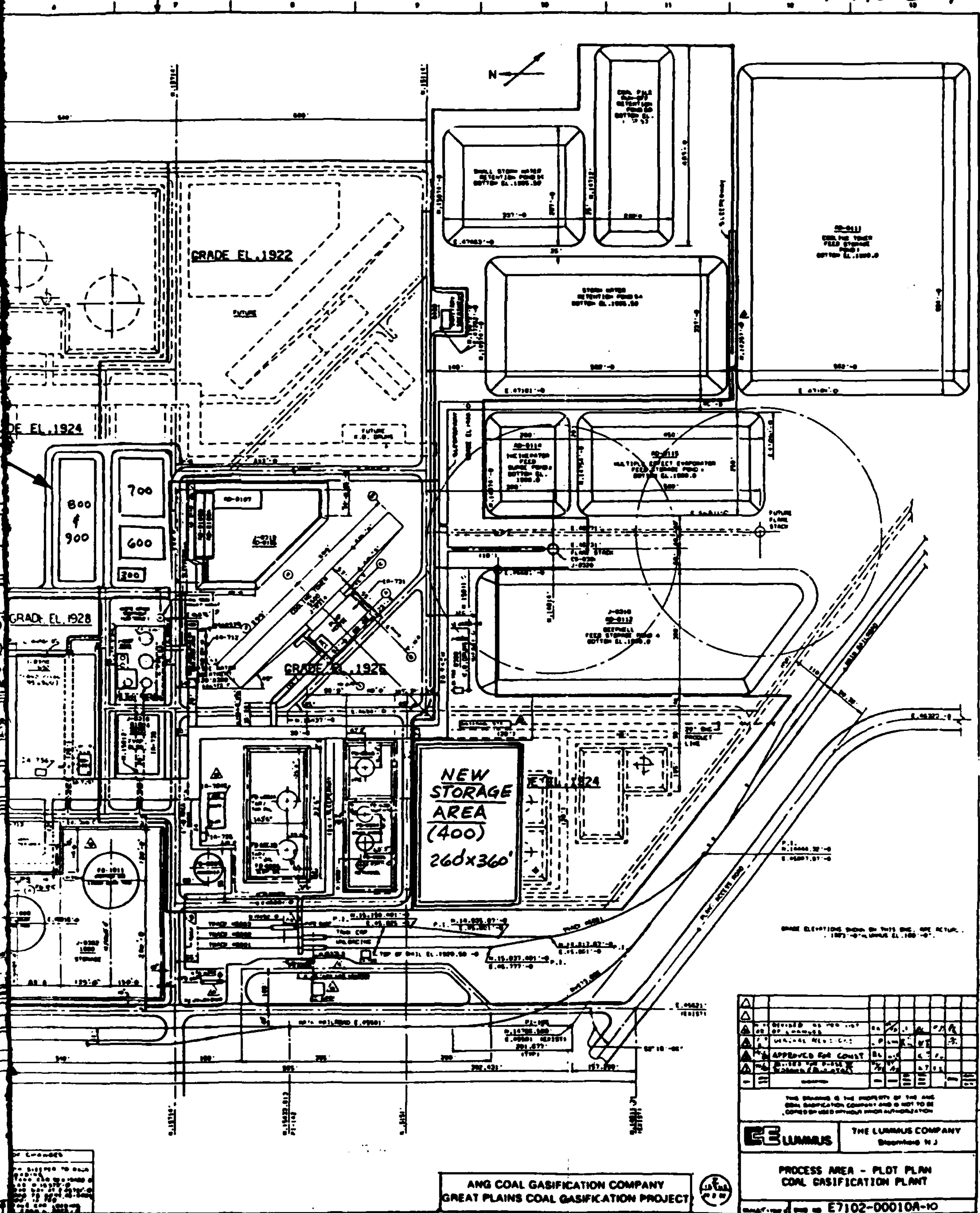
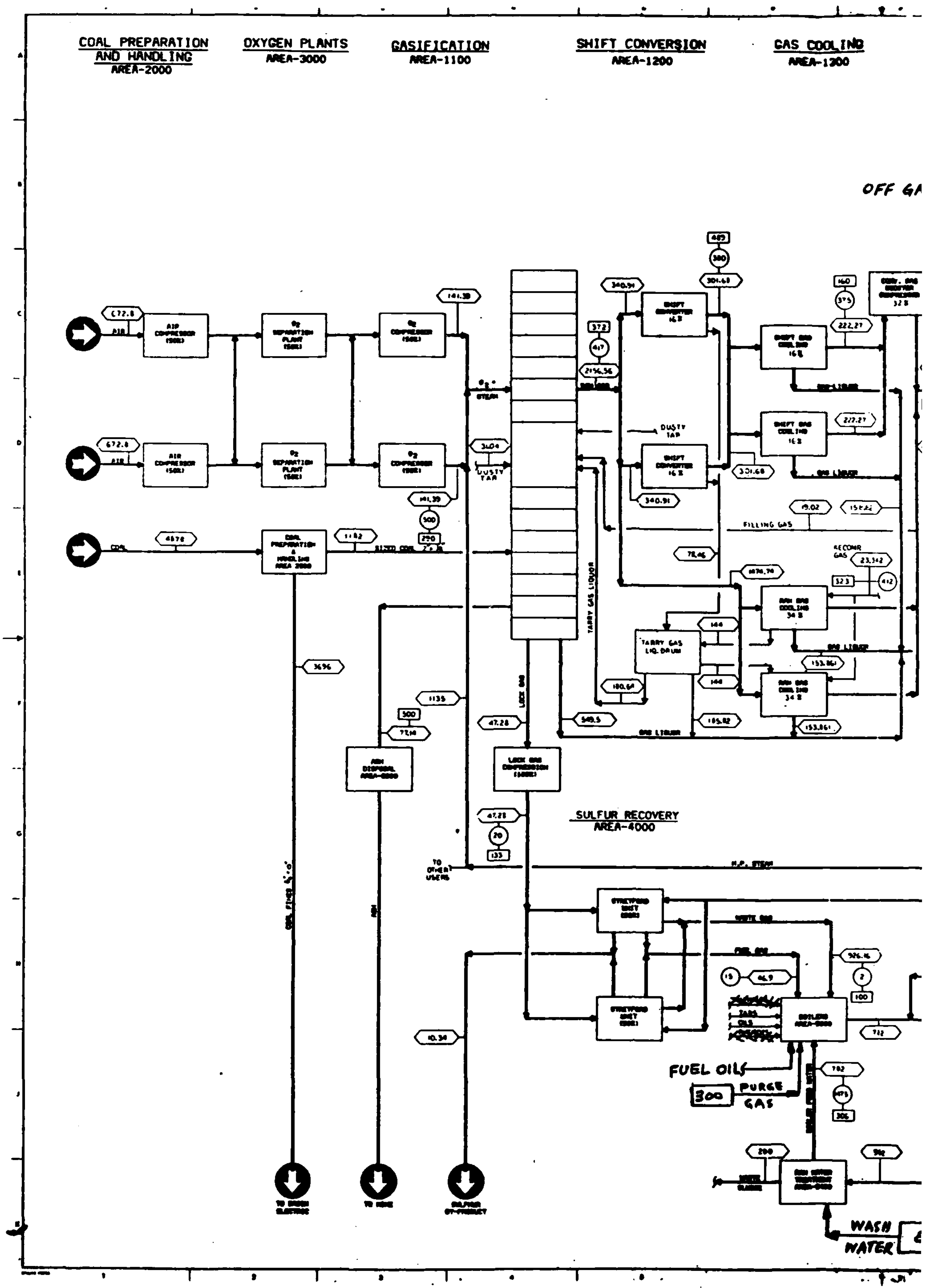
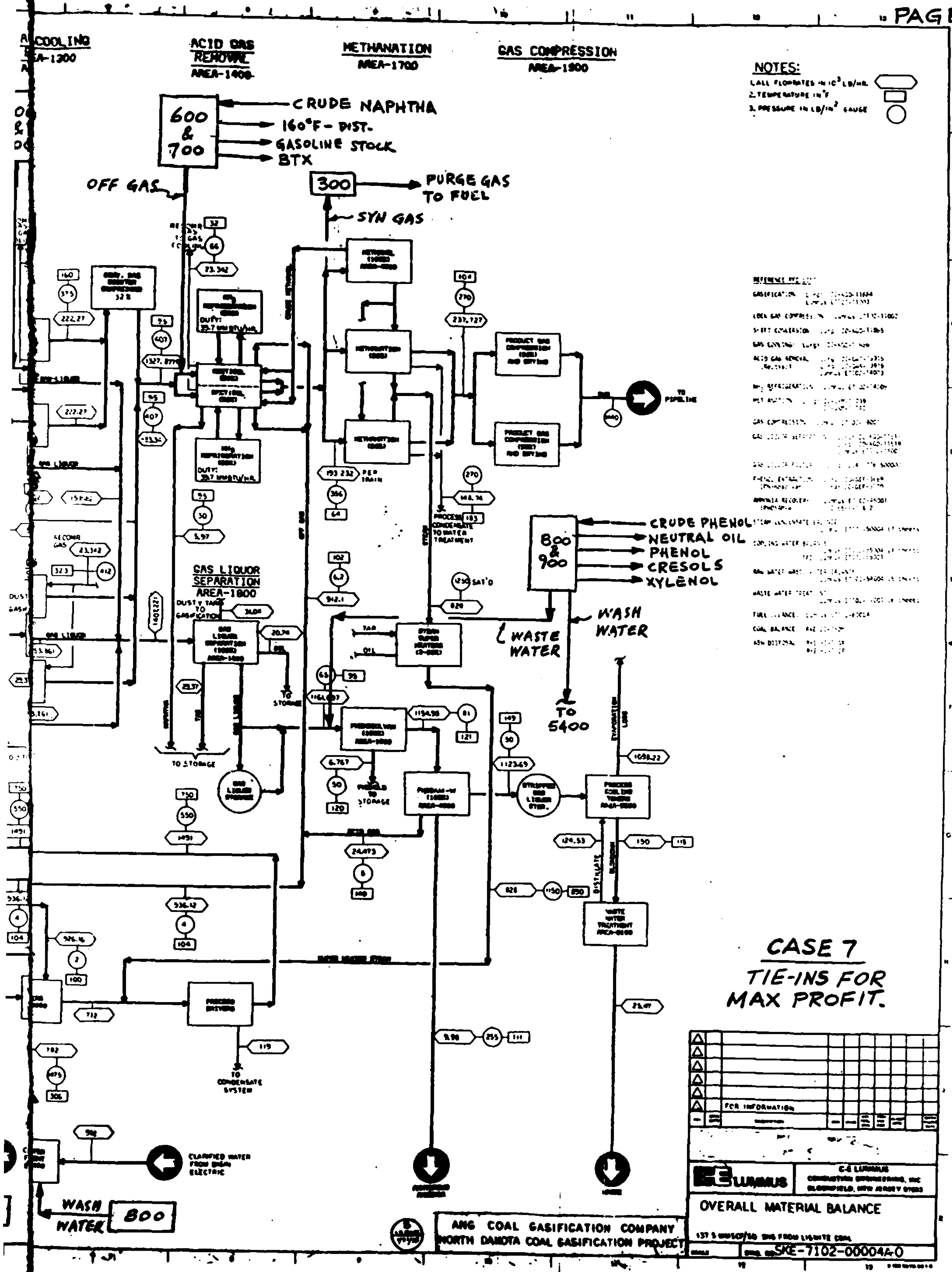
GRADE EL.1926
FUTURE
GRADE EL.1924
GRADE EL.1932
GRADE EL.1930
GRADE EL.1928


GRADE EL.1922
GRADE EL.1924
GRADE EL.1928
GRADE EL.1926
EL.1924
700
800
900
600
200
NEW
STORAGE
AREA
(400)
260x360'
N
THE LUMMUS COMPANY
LUMMUS
PROCESS AREA - PLOT PLAN
COAL GASIFICATION PLANT
ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT
E7102-00010A-10

CASE 7



COAL PREPARATION AND HANDLING
AREA-2000
OXYGEN PLANTS
AREA-3000
GASIFICATION
AREA-1100
SHIFT CONVERSION
AREA-1200
GAS COOLING
AREA-1200
SULFUR RECOVERY
AREA-4000
AIR
COAL
TARRY GAS LIQUOR
DUSTY TAR
FILLING GAS
RECOMP GAS
TARRY GAS LIQ. DRUM
LOCK GAS
TO OTHER USERS
FUEL OILS
PURGE GAS
WASH WATER


ACID GAS REMOVAL
METHANATION
AREA-1700
GAS COMPRESSION
AREA-1900
NOTES:
600 & 700
CRUDE NAPHTHA
160°F - DIST.
GASOLINE STOCK
BTX
OFF GAS
300
PURGE GAS TO FUEL
SYN GAS
GAS LIQUOR SEPARATION
AREA-1800
800 & 900
CRUDE PHENOL
NEUTRAL OIL
PHENOL
CRESOLS
XYLENOL
WASH WATER
WASTE WATER
TO 5400
TO STORAGE
TO CONDENSATE SYSTEM
CLARIFIED WATER FROM BASIN ELECTRIC
WASH WATER
800
CASE 7
TIE-INS FOR MAX PROFIT.
FOR INFORMATION
C-E LUMMUS
OVERALL MATERIAL BALANCE
ANG COAL GASIFICATION COMPANY
NORTH DAKOTA COAL GASIFICATION PROJECT
SKE-7102-00004A-0

CASE 7

APPENDIX A
AREA 800/900
PAGE 1

CRUDE PHENOLS PRODUCTION 113 MILLION LB/YR CRUDE PHENOL COLUMN
STREAM FACTOR .89
ON-STREAM HR/YR 7796.4

STREAM NO				1		13				
STREAM DESCRIPTION				CRUDE PHENOL	PHENOL RECYCLE	o-CRESOL PHENOL	FEED	TOTAL OVERHEAD	OVERHEAD PHENOL	WATER + LIGHTS
COMPONENT		M.P. DEG C	B.P. DEG C	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
WATER	H2O		100	724	0		724	724	55	670
METHANOL				0	0					
HEXANE				0	0					
SURFUIC ACID	H2SO4			0	0					
LIGHTS				290	1		291	291		291
PYRIDINES	C5H5N		115	145	0		145	145	1	144
PHENOL	C6H5OH	43	181	4925	490	703	6118	5479	5426	53
NEUTRAL OIL				579	1		580	82	82	
o-CRESOL	CH3C6H4OH 1,2	30	191	869	0	7	877	0		
p-CRESOL	CH3C6H4OH 1,4	35.5	202	0	0	0	0	0		
m-CRESOL	CH3C6H4OH 1,3	11	202	1977	0		1978	0		
GUAIACOL	HOC6H4OCH3 1,2	32	205	0	0		0	0		
o-ETHYL PHENOL	OHC6H4C2H5		207	58	0		58	0		
2,4-XYLENOL	HOC6H3(CH3)2	26	211	1050	0		1050	0		
2,5-XYLENOL	HOC6H3(CH3)2	75	212	0	0		0	0		
2,6-XYLENOL	HOC6H3(CH3)2		212	0	0		0	0		
p-ETHYL PHENOL	OHC6H4C2H5		214	0	0		0	0		
2,3-XYLENOL	HOC6H3(CH3)2		218	0	0		0	0		
3,5-XYLENOL	HOC6H3(CH3)2	68	219	0	0		0	0		
m-ETHYL PHENOL	OHC6H4C2H5		219	109	0		109	0		
3,4-XYLENOL	HOC6H3(CH3)2	62.5	225	0	0		0	0		
CATECHOL	C6H4(OH)2 1,2	105	245	0	0		0	0		
RESIDUE				3767	0		3767	0		
RESORCINOL	C6H4(OH)2 1,3	110	281	0	0		0	0		
UNKNOWNS				0	0		0	0		
TOTAL #/HR				14494	492	710	15697	6721	5564	1157
API				1.7	1.2	1.2	1.7	3.3	1.5	12.4
S.G.				1.062	1.066	1.066	1.062	1.05	1.064	0.983
#/GAL				8.857	8.890	8.890	8.858	8.756	8.873	8.197
BSD				935	32	46	1013	439	358	81
BCD				832	28	41	901	390	319	72

WTX / RECOVERY	CRUDE PHENOL WT %	TOTAL OVERHEAD WT %	OVERHEAD PHENOL WT %	WATER + LIGHTS WT %
WATER	5.00	100	1	17.68
METHANOL				
HEXANE				
SURFUIC ACID				
LIGHTS	2.00	100		
PYRIDINES	1.00	100	0.02	
PHENOL	34.00	89.55		1.40
NEUTRAL OIL	4.00	1.50	1.5	
o-CRESOL	6.00			
p-CRESOL	0.00			
m-CRESOL	13.65			
GUAIACOL				
o-ETHYL PHENOL	0.40			
2,4-XYLENOL	7.25			
2,5-XYLENOL				
2,6-XYLENOL				
p-ETHYL PHENOL				
2,3-XYLENOL				
3,5-XYLENOL				
m-ETHYL PHENOL	0.75			
3,4-XYLENOL				
CATECHOL				
RESIDUE	26.00			
RESORCINOL				
UNKNOWNS				

				LIGHT ENDS COLUMN					PHENOL COLUMN		
STREAM NO		2									5
STREAM DESCRIPTION / COMPONENT	OVERHEAD WATER #/HR	OVERHEAD LIGHTS #/HR	BOTTOMS #/HR	FEED #/HR	WATER #/HR	OVERHEAD LIGHTS #/HR	OVERHEAD WATER #/HR	BOTTOMS #/HR	TOTAL OVERHEAD PHENOL #/HR	BOTTOMS #/HR	PHENOL PRODUCT #/HR
WATER	670		0	55	4022	0	4077	0			0
METHANOL				0		0	0	0			0
HEXANE				0		0	0	0			0
SURFUIC ACID				0		0	0	0			0
LIGHTS		291	0	291		280	0	11	11		10
PYRIDINES		144	0	145		145	0	0			0
PHENOL	18	35	639	5461	318	0	322	5456	5386	70	4896
NEUTRAL OIL			498	82		1	0	81	11	70	10
o-CRESOL			877	0		0	0	0	5		5
p-CRESOL			0	0		0	0	0	0		0
m-CRESOL			1978	0		0	0	0	5		5
GUAIACOL			0	0		0	0	0	0		0
o-ETHYL PHENOL			58	0		0	0		0		0
2,4-XYLENOL			1050	0		0	0		0		0
2,5-XYLENOL			0	0		0	0		0		0
2,6-XYLENOL			0	0		0	0		0		0
p-ETHYL PHENOL			0	0		0	0		0		0
2,3-XYLENOL			0	0		0	0		0		0
3,5-XYLENOL			0	0		0	0		0		0
m-ETHYL PHENOL			109	0		0	0		0		0
3,4-XYLENOL			0	0		0	0		0		0
CATECHOL			0	0		0	0		0		0
RESIDUE			3767	0		0	0		0		0
RESORCINOL			0	0		0	0		0		0
UNKNOWNS			0	0		0	0		0		0
TOTAL #/HR	687	469	8976	6033	4340	425	4399	5549	5419	141	4926
API	10.0	15.1	5.9	2.6	10.0	14.4	10.0	2.0	1.2	10.0	1.2
S.G.	1	0.965	1.03	1.055	1.000	0.970	1.000	1.060	1.066	1.000	1.066
#/GAL	8.339	8.047	8.589	8.802	8.343	8.093	8.343	8.843	8.894	8.343	8.89
BSD	47	33	597	392	297	30	301	359	348	10	316
BCD	42	30	531	349	265	27	268	319	310	9	282
WT% / RECOVERY					WT %	WT %	WT %	WT %	WT %	WT %	WT %
WATER					66.67		100				91
METHANOL											91
HEXANE											91
SURFUIC ACID											91
LIGHTS								0.2	0.2		91
PYRIDINES						100	0				91
PHENOL					8		8	99.8	99.4	50	91
NEUTRAL OIL						1	0		0.2	50	91
o-CRESOL									0.1		91
p-CRESOL											91
m-CRESOL									0.1		91
GUAIACOL											91
o-ETHYL PHENOL											91
2,4-XYLENOL											91
2,5-XYLENOL											91
2,6-XYLENOL											91
p-ETHYL PHENOL											91
2,3-XYLENOL											91
3,5-XYLENOL											91
m-ETHYL PHENOL											91
3,4-XYLENOL											91
CATECHOL											91
RESIDUE											91
RESORCINOL											91
UNKNOWNS											91

		ACID FLASH					TAR WASHING			EXTRACTOR COLUMN	
STREAM NO					7		4		8		
STREAM DESCRIPTION	PHENOL RECYCLE	FEED	SULFURIC ACID	TOTAL FEED	OVERHEAD	TAR OIL	WASH WATER	WASTE WATER	WASHED TAR	FEED FR ACID FLASH	HEXANE SOLVENT
COMPONENT	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
WATER	0	0	6	6	6	0	1246	1246	0	6	0
METHANOL	0				0				0	0	0
HEXANE	0				0				0	0	11372
SURFUIC ACID	0		285	285	0	285		285	0	0	0
LIGHTS	1	0		0	0	0			0	0	0
PYRIDINES	0	0		0	0	0			0	0	0
PHENOL	490	710		710	710	0			0	710	0
NEUTRAL OIL	1	569		569	569	0			0	569	0
o-CRESOL	0	877		877	877	0			0	877	0
p-CRESOL	0	0		0	0	0			0	0	0
m-CRESOL	0	1978		1978	1938	40			40	1938	0
GUAIACOL	0	0		0	0	0			0	0	0
o-ETHYL PHENOL	0	58		58	57	1			1	57	0
2,4-XYLENOL	0	1050		1050	1029	21			21	1029	0
2,5-XYLENOL	0	0		0	0	0			0	0	0
2,6-XYLENOL	0	0		0	0	0			0	0	0
p-ETHYL PHENOL	0	0		0	0	0			0	0	0
2,3-XYLENOL	0	0		0	0	0			0	0	0
3,5-XYLENOL	0	0		0	0	0			0	0	0
m-ETHYL PHENOL	0	109		109	106	2			2	106	0
3,4-XYLENOL	0	0		0	0	0			0	0	0
CATECHOL	0	0		0	0	0			0	0	0
RESIDUE	0	3767		3767	377	3390			3390	377	0
RESORCINOL	0	0		0	0	0			0	0	0
UNKNOWNS	0	0		0	0	0			0	0	0
TOTAL #/HR	493	9117	291	9408	5669	3739	1246	1532	3454	5669	11372
API	1.2	6.0		4.1	11.9					11.9	
S.G.	1.066	1.029	1.83	1.043	0.987	1.140	1.000	1.100	1.100	0.987	0.660
#/GAL	8.89	8.585	15.27	8.703	8.23	9.51	8.34	9.18	9.18	8.23	5.51
BSD	32	607	11	618	393	225	85	95	215	393	1180
BCD	28	540	10	550	350	200	76	85	191	350	1050

WTX / RECOVERY	WT %		WT %		WT %		WTX				WT %
WATER	9		2		100		33.33				
METHANOL	9										
HEXANE	9										100
SURFUIC ACID	9		3								
LIGHTS	9				100						
PYRIDINES	9				100						
PHENOL	9				100						
NEUTRAL OIL	9				100						
o-CRESOL	9				100						
p-CRESOL	9				98						
m-CRESOL	9				98						
GUAIACOL	9				98						
o-ETHYL PHENOL	9				98						
2,4-XYLENOL	9				98						
2,5-XYLENOL	9				98						
2,6-XYLENOL	9				98						
p-ETHYL PHENOL	9				98						
2,3-XYLENOL	9				98						
3,5-XYLENOL	9				98						
m-ETHYL PHENOL	9				98						
3,4-XYLENOL	9				98						
CATECHOL	9				95						
RESIDUE	9				10						
RESORCINOL	9				5						
UNKNOWNS	9				5						

				HEXANE COLUMN		METHANOL COLUMN		DRYING COLUMN		o-CRESOL COLUMN	
STREAM NO					10				11	13	12
STREAM DESCRIPTION	METHANOL/ WATER SOLVENT	OVERHEAD	BOTTOMS	HEXANE OVERHEAD	NEUTRAL OIL BOTTOMS	METHANOL/ WATER OVERHEAD	WET CRESYLIC ACID	DRYING COLUMN OVHD	DRY CRESYLIC ACID	o-CRESOL PHENOL	o-CRESOL
COMPONENT	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
WATER	1765	0	1771	0	0	1594	177	177	0		
METHANOL	3278	0	3278	0	0	3278	0				
HEXANE	0	11372	0	11372	0	0	0				
SURFUIC ACID	0	0	0	0	0	0	0		0		
LIGHTS	0	0	0	0	0	0	0		0		
PYRIDINES	0	0	0	0	0	0	0		0		
PHENOL	0	0	710	0	0	0	710		710	703	7
NEUTRAL OIL	0	569	0	0	569	0	0		0		
o-CRESOL	0	0	877	0	0	0	877		877	7	826
p-CRESOL	0	0	0	0	0	0	0		0	0	0
m-CRESOL	0	0	1938	0	0	0	1938		1938		17
GUAIACOL	0	0	0	0	0	0	0		0		
o-ETHYL PHENOL	0	0	57	0	0	0	57		57		
2,4-XYLENOL	0	0	1029	0	0	0	1029		1029		
2,5-XYLENOL	0	0	0	0	0	0	0		0		
2,6-XYLENOL	0	0	0	0	0	0	0		0		
p-ETHYL PHENOL	0	0	0	0	0	0	0		0		
2,3-XYLENOL	0	0	0	0	0	0	0		0		
3,5-XYLENOL	0	0	0	0	0	0	0		0		
m-ETHYL PHENOL	0	0	106	0	0	0	106		106		
3,4-XYLENOL	0	0	0	0	0	0	0		0		
CATECHOL	0	0	0	0	0	0	0		0		
RESIDUE	0	0	377	0	0	0	377		377		
RESORCINOL	0	0	0	0	0	0	0		0		
UNKNOWNS	0	0	0	0	0	0	0		0		
TOTAL #/HR	5043	11940	10142	11372	569	4871	5271	177	5094	710	849
API							8.6	8.6	8.6	2	5.9
S.G.	0.878	0.669	0.950	0.660	0.914	0.876	1.030	1.000	1.030	1.066	1.035
#/GAL	7.33	5.58	7.93	5.51	7.63	7.31	8.59	8.34	8.59	8.89	8.63
BSD	393	1222	731	1180	43	381	351	12	339	46	56
BCD	350	1088	651	1050	38	339	312	11	301	41	50
WT% / RECOVERY	WT %	RECOVERY	RECOVERY							RECOVERY	RECOVERY
WATER	35		100			90	10				
METHANOL	65		100			100					
HEXANE		100		100							
SURFUIC ACID											
LIGHTS											
PYRIDINES											
PHENOL			100							99	1
NEUTRAL OIL		100									
o-CRESOL			100							1	95
p-CRESOL											
m-CRESOL											2
GUAIACOL											
o-ETHYL PHENOL											
2,4-XYLENOL											
2,5-XYLENOL											
2,6-XYLENOL											
p-ETHYL PHENOL											
2,3-XYLENOL											
3,5-XYLENOL											
m-ETHYL PHENOL											
3,4-XYLENOL											
CATECHOL											
RESIDUE											
RESORCINOL											
UNKNOWNS											

	m,p-CRESOL COLUMN				XYLENOL COLUMN			
STREAM NO		14			15	16		6
STREAM DESCRIPTION	o-CRESOL COLUMN BOTTOMS	m/pCRESOL	o-ETHYL PHENOL	m/pCRESOL COLUMN BOTTOMS	XYLENOLS	HEAVIES	PHEN EXT NEUTRAL OIL	TOTAL NEUTRAL OIL
COMPONENT	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
WATER	0	0		0	0	0	0	0
METHANOL							0	0
HEXANE							0	0
SURFUIC ACID	0	0		0	0	0	0	0
LIGHTS	0	0		0	0	0	280	280
PYRIDINES	0	0		0	0	0	145	145
PHENOL	0	0		0	0	0	0	0
NEUTRAL OIL	0	0		0	0	0	570	570
o-CRESOL	44	44		0	0	0	0	0
p-CRESOL	0	0		0	0	0	0	0
m-CRESOL	1922	1826	0	96	96	0	40	40
GUAIACOL	0			0	0	0	0	0
o-ETHYL PHENOL	57	54	0	3	3	0	1	1
2,4-XYLENOL	1029	51		978	929	49	21	70
2,5-XYLENOL	0			0		0	0	0
2,6-XYLENOL	0			0		0	0	0
p-ETHYL PHENOL	0			0		0	0	0
2,3-XYLENOL	0			0		0	0	0
3,5-XYLENOL	0			0		0	0	0
m-ETHYL PHENOL	106			106	49	58	2	60
3,4-XYLENOL	0			0		0	0	0
CATECHOL	0			0		0	0	0
RESIDUE	377			377	0	377	3390	3767
RESORCINOL	0			0		0	0	0
UNKNOWNS	0			0		0	0	0
TOTAL #/HR	3535	1975	0	1560	1077	483	4448	4931
API	5.9	5.9	5.9	14.4	13.6	-0.5	2.1	0.7
S.G.	1.024	1.034	1.030	1.010	0.975	1.08	1.059	1.07
#/GAL	8.54	8.63	8.59	8.43	8.13	9.01	8.834	8.93
BSD	237	131	0	106	76	31	288	316
BCD	210	116	0	94	67	27	256	281

WT% / RECOVERY		RECOVERY	RECOVERY		RECOVERY			
WATER								
METHANOL								
HEXANE								
SURFUIC ACID								
LIGHTS								
PYRIDINES								
PHENOL								
NEUTRAL OIL								
o-CRESOL								
p-CRESOL								
m-CRESOL		95						
GUAIACOL								
o-ETHYL PHENOL		95	0					
2,4-XYLENOL		5			95			
2,5-XYLENOL								
2,6-XYLENOL								
p-ETHYL PHENOL								
2,3-XYLENOL								
3,5-XYLENOL								
m-ETHYL PHENOL					5			
3,4-XYLENOL								
CATECHOL								
RESIDUE					0			
RESORCINOL								
UNKNOWNS								

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SIMULATION SCIENCES INC. PROCESS™ PAGE 15
PROJECT GP JET FUELS ML
PROBLEM NAPHDT SOLUTION DEC 31 1987

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CASE 7
AREA 600

NAPHTHA FRACTIONATION

STREAM ID.	1	2	3	4
STREAM NAME.........	RECTISOL NAP	160 MINUS	160 PLUS	DIST OFFGAS
STREAM PHASE........	LIQUID	LIQUID	LIQUID	VAPOR
TEMPERATURE, DEG F	33.0000	100.0000	272.4216	100.0000
PRESSURE, PSIA	46.1785	40.0000	47.9999	40.0000
RATE LB MOLS/HR	114.1908	35.9744	78.2163	
RATE LB /HR	8737.9805	2168.0786	6570.0576	NORMALLY
ENTHALPY MM BTU /HR	-0.0349	0.0386	0.6911	NO
ENTHALPY BTU /LB	-3.9967	17.8333	105.1917	FLOW
MOLECULAR WEIGHT....	76.5209	60.2673	83.9986	
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	8737.9805	2168.0786	6570.0576	0.0000
ACT.RATE BBL/DAY	713.18	208.33	607.62	0.00
STD. LV RATE BBL/HR	30.20	8.42	21.77	0.00
CP, BTU /LB F	0.4193	0.5621	0.4886	0.0000
MOLECULAR WEIGHT....	76.5209	60.2673	83.9986	0.0000
ACT.DENS LB /FT3	52.3723	44.4841	46.2197	0.0000
STD. API GRAVITY....	39.5750	60.6584	32.5314	0.0000
*** DRY BASIS ***				
RATE LB /HR	8693.9824	2124.0786	6570.0576	0.0000
MOLECULAR WEIGHT....	77.7996	63.3448	83.9986	0.0000
UOP K	10.4363	11.4820	10.0981	0.0000
FLASH POINT, DEG F	-11.2329	-33.4546	19.7649	0.0000
CRIT. TEMP, F	530.7584	449.4239	565.6373	0.0000
CRIT. PRES, PSIA	665.3411	682.4471	658.0043	0.0000
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	8693.9824	2124.0786	6570.0576	0.0000
ACT.RATE BBL/DAY	710.17	205.39	607.62	0.00
CP, BTU /LB F	0.4169	0.5530	0.4886	0.0000
MOLECULAR WEIGHT....	77.7996	63.3448	83.9986	0.0000
ACT.DENS LB /FT3	52.3297	44.2255	46.2197	0.0000
STD. API GRAVITY....	39.7443	61.7552	32.5314	0.0000
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]

APPENDIX B

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	5	6	7	8	9	10	11
STREAM NAME..........	MAKE-UP GAS	REACTOR INLE	REACTOR EFFL	WASH WATER	COLD SEP LIQ	PURGE GAS	RECYCLE GAS
STREAM PHASE........	VAPOR	VAPOR	VAPOR	LIQUID	LIQUID	VAPOR	VAPOR
TEMPERATURE, DEG F	70.0000	425.0000	475.0000	100.0000	120.0000	120.0000	120.0000
PRESSURE, PSIA	340.0000	800.0000	775.0000	735.0000	710.0000	710.0000	710.0000
RATE LB MOLS/HR	31.0000	277.6176	264.3793	138.7732	82.3024	5.3009	168.4035
RATE LB /HR	62.5395	7324.1631	7321.4014	2499.9995	6446.4883	21.7711	691.6453
ENTHALPY MM BTU /HR	-0.0644	2.2362	2.5851	0.1750	0.2584	-0.0062	-0.1984
ENTHALPY BTU /LB	-1030.5195	305.3162	353.0820	69.9971	40.0906	-286.8285	-286.8286
MOLECULAR WEIGHT....	2.0174	26.3822	27.6928	18.0150	78.3268	4.1071	4.1071
*** VAPOR PHASE ***							
RATE LB /HR	62.5395	7324.1631	7321.4014	0.0000	0.0000	21.7711	691.6453
ACT.RATE FT3/SEC	0.15	0.89	0.93	0.00	0.00	0.01	0.42
STD.RATE MM FT3/DAY	0.28	2.53	2.41	0.00	0.00	0.05	1.53
CP, BTU /LB F	3.4201	0.6138	0.6083	0.0000	0.0000	1.8091	1.8091
MOLECULAR WEIGHT....	2.0174	26.3822	27.6928	0.0000	0.0000	4.1071	4.1071
ACT.DENS LB /FT3	0.1189	2.2787	2.1894	0.0000	0.0000	0.4569	0.4569
COMPRESSIBILITY (Z).	1.0145	0.9757	0.9773	0.0000	0.0000	1.0260	1.0260
*** LIQUID PHASE ***							
RATE LB /HR	0.0000	0.0000	0.0000	2499.9995	6446.4883	0.0000	0.0000
ACT.RATE BBL/DAY	0.00	0.00	0.00	172.41	541.70	0.00	0.00
STD. LV RATE BBL/HR	0.00	0.00	0.00	7.15	21.83	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.9941	0.4159	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	18.0150	78.3268	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	61.9844	50.8694	0.0000	0.0000
STD. API GRAVITY....	0.0000	0.0000	0.0000	10.0635	36.1255	0.0000	0.0000
*** DRY BASIS ***							
RATE LB /HR	0.0000	0.0000	0.0000	0.0000	6443.5264	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000	78.4475	0.0000	0.0000
UOP K	0.0000	0.0000	0.0000	0.0000	10.1772	0.0000	0.0000
FLASH POINT, DEG F	0.0000	0.0000	0.0000	0.0000	-75.8712	0.0000	0.0000
CRIT. TEMP, F	0.0000	0.0000	0.0000	0.0000	519.0099	0.0000	0.0000
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	0.0000	678.9440	0.0000	0.0000
*** VAPOR PHASE ***							
RATE LB /HR	62.5395	7316.9844	7187.2246	0.0000	0.0000	21.5452	684.4674
ACT.RATE FT3/SEC	0.15	0.89	0.90	0.00	0.00	0.01	0.42
STD.RATE MM FT3/HR	0.01	0.11	0.10	0.00	0.00	0.00	0.06
CP, BTU /LB F	3.4201	0.6139	0.6113	0.0000	0.0000	1.8233	1.8233
MOLECULAR WEIGHT....	2.0174	26.3942	27.9733	0.0000	0.0000	4.0741	4.0741
ACT.DENS LB /FT3	0.1189	2.2798	2.2128	0.0000	0.0000	0.4532	0.4532
COMPRESSIBILITY (Z).	1.0145	0.9736	0.9768	0.0000	0.0000	1.0261	1.0261
VISCOSITY, CP	0.0076	0.0185	0.0191	0.0000	0.0000	0.0106	0.0106
*** LIQUID PHASE ***							
RATE LB /HR	0.0000	0.0000	0.0000	0.0000	6443.5264	0.0000	0.0000
ACT.RATE BBL/DAY	0.00	0.00	0.00	0.00	541.49	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000	0.4157	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000	78.4475	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.0000	0.0000	50.8653	0.0000	0.0000
STD. API GRAVITY....	0.0000	0.0000	0.0000	0.0000	36.1375	0.0000	0.0000

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NAPHTHA HDT PRODUCT STRIPPER

STREAM ID.	12	13	15	17
STREAM NAME..........	SOUR H2O	STAB PRD	STAB OFFGAS	STAB SOUR H2O
STREAM PHASE........	LIQUID	LIQUID	VAPOR	LIQUID
TEMPERATURE, DEG F	120.0000	287.5287	100.0000	100.0000
PRESSURE, PSIA	710.0000	70.0000	60.0000	60.0000
RATE LB MOLS/HR	147.1456	73.8282	8.4432	0.0309
RATE LB /HR	2661.4951	6135.1143	313.3105	0.5570
ENTHALPY MM BTU /HR	0.0780	0.6899	0.0653	0.0000
ENTHALPY BTU /LB	29.3215	112.4480	208.5772	67.9931
MOLECULAR WEIGHT....	18.0875	83.0998	37.1079	18.0150
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	313.3105	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.23	0.00
STD.RATE MM FT3/DAY	0.00	0.00	0.08	0.00
CP, BTU /LB F	0.0000	0.0000	0.3973	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	37.1079	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.3844	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9645	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	2661.4951	6135.1143	0.0000	0.5570
ACT.RATE BBL/DAY	185.82	576.27	0.00	0.04
STD. LV RATE BBL/HR	7.66	20.33	0.00	0.00
CP, BTU /LB F	1.1616	0.4890	0.0000	0.9977
MOLECULAR WEIGHT....	18.0875	83.0998	0.0000	18.0150
ACT.DENS LB /FT3	61.2244	45.5081	0.0000	61.9843
STD. API GRAVITY....	10.8946	32.5122	0.0000	10.0635
*** DRY BASIS ***				
RATE LB /HR	37.6836	6135.1016	0.0000	0.0000
MOLECULAR WEIGHT....	25.1254	83.1004	0.0000	0.0000
UOP K	10.2499	10.0651	0.0000	0.0000
FLASH POINT, DEG F	-5.3046	0.7399	0.0000	0.0000
CRIT. TEMP, F	214.9373	561.6288	0.0000	0.0000
CRIT. PRES, PSIA	1412.3528	667.3295	0.0000	0.0000
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	310.9178	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.22	0.00
STD.RATE MM FT3/HR	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.3971	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	37.4130	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.3877	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9640	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0102	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	37.6836	6135.1016	0.0000	0.0000
ACT.RATE BBL/DAY	4.07	576.27	0.00	0.00
CP, BTU /LB F	0.8241	0.4890	0.0000	0.0000
MOLECULAR WEIGHT....	25.1254	83.1004	0.0000	0.0000
ACT.DENS LB /FT3	39.6032	45.5080	0.0000	0.0000
STD. API GRAVITY....	[illegible].7618	32.5122	0.0000	0.0000
VISCOSITY, CP	[illegible]	[illegible]	[illegible]	[illegible]